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Los Angeles County Sanitation District, Whittier, CA

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16. ABSTRACT

Two 6.3 l/sec (0.15 mgd), two-stage, packed-bed, downflow granular activated carbon pilot plants were operated continuously for 33 months using unfiltered and unchlorinated activated sludge plant effluent. The main objective of the study was to compare the performance of granular carbons of different particle size. The data collected during this study has demonstrated the efficacy of the two stage carbon adsorptive system in consistently producing effluent of excellent overall quality. Effluent averaged 6-7 mg/l DCOD and 2 mg/l S.S.

The carbon capacity with the 8x30 mesh carbon decreased about 25% after four adsorption cycles, resulting in an apparent steady state capacity of 0.26 lbs. DCOD removed/lb carbon. A 23% decrease in carbon capacity occurred after three adsorption cycles with the 12x40 mesh carbon. The 12x40 carbon has about 13% more DCOD removal capacity than the 8x30 carbon. While the smaller carbon showed slightly higher treatment capacity than the larger carbon, the latter has lower initial cost, lower pressure loss and lower regeneration loss. Thus the larger size carbon was more economical,

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TWO-STAGE GRANULAR ACTIVATED CARBON TREATMENT

by

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CINCINNATI, OHIO 45268

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

One method of tertiary or advanced treatment beyond the conventional secondary treatment level is granular carbon adsorption. The treatment process results in significant additional reduction in suspended solids and organics. Original systems employed several stages of contact to provide optimum utilization of carbon adsorption capacity. In this study an evaluation of a two-stage rather than a multi-stage system was conducted to ascertain the performance of this lower capital cost version of the process. In addition a comparison of performance with two different granular carbon sizes was conducted.

Francis T. Mayo, Director
Municipal Environmental
Research Laboratory

ABSTRACT

Two 6.3 l/sec (0.15 mgd), two-stage, packed-bed, downflow granular activated carbon pilot plants were operated continuously for 33 months using unfiltered and unchlorinated activated sludge plant effluent. The main objective of the study was to evaluate the effect of repeated thermal regeneration cycles on the adsorption capacity, regeneration loss and pressure drop buildup of carbon with different particle size. Performance data, collected during the field study, has demonstrated the stability of the two-stage carbon adsorptive system in consistently producing effluents of excellent overall quality.

The carbon capacity in the Filtrasorb 300 system (8x30 mesh carbon) decreased about 25% after four adsorption cycles, resulting in an apparent steady state capacity of 0.26 lbs. DCOD removed/lb. carbon. A 23% decrease in carbon capacity occurred after three adsorption cycles in the Filtrasorb 400 system, (12x40 mesh carbon). The 400 system has about 13% more DCOD removal capacity than the 300 system. While the 400 system showed slightly higher carbon capacity than the 300 system, the latter has the advantage of not only lower initial cost, but also lower in both pressure loss and regeneration loss.

The estimated total treatment cost for a 0.44 m^3/sec (10 mgd), two-stage carbon adsorption system with 8x30 mesh carbon for treating an activated sludge plant effluent is 11.52¢/1000 gallons.

This report was submitted by the Los Angeles County Sanitation Districts, Los Angeles, California, in fulfillment of Contract Number 14-12-150 under the sponsorship of the Environmental Protection Agency. This report covers the period March, 1970, to December, 1973, and work was completed as of December, 1975.

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The untiring efforts and assistance of both the laboratory and the pilot plant operating personnel of the Pomona Advanced Waste Treatment Research Facility are gratefully acknowledged.

SECTION I

INTRODUCTION

The use of activated carbon for the removal of organic contaminants has been known for a long time and is well decumented in literature (1). However, it is only in recent years that research efforts have been directed towards the application of activated carbon for the removal of residual organic matter from wastewater.

Since 1965, the Environmental Protection Agency and the Los Angeles County Sanitation Districts have jointly operated a 12.6 l/sec (0.3 mgd) activated carbon adsorption pilot plant at the Sanitation District's Advanced Waste Treatment Research Facility at Pomona, California. Between 1965 and 1969, the 12.6 l/sec (0.3 mgd) pilot plant was operated as a 4-stage adsorption system with the primary objective of establishing the technical and economic feasibility of using granular activated carbon for removing soluble organics from secondary effluent and to obtain cost data and operating experience on carbon regeneration. The results of the 4-year field study demonstrated successfully the effectiveness of granular activated carbon in the adsorption of residual organic matter from an activated sludge plant effluent. In addition, extensive operating data collected over the years have demonstrated that thermal regeneration of exhausted carbon is an economically feasible process. Thus, based on the experience with the 4-stage adsorption study and following an appraisal of the state-of-the-art of granular carbon contacting systems published in 1969 (2), the carbon columns were modified in March 1970 to operate as two parallel 6.3 l/sec (0.15 mgd) two-stage adsorption systems.

The objectives of the two-stage adsorption-system study were to determine on a long-term basis the effect of carbon particle size on adsorption rate and adsorption capacity at 40 minute contact time; and to evaluate the effect of repeated thermal regeneration cycles on the adsorption capacity, regeneration loss and pressure drop build-up of carbon with different particle size. The two parallel two-stage adsorption systems, one containing a 12 X 40 mesh carbon similar to that used in the 4-stage adsorption system and the other an 8 X 30 mesh carbon, were operated in a similar fashion and at the same contact time as the 4-stage system.

The data presented in this final report are the results of four adsorption cycles for the 8 X 30 mesh carbon, which required 33 months to complete and 3 adsorption cycles covering 26 months for the 12 X 40 mesh carbon.

SECTION II

CONCLUSIONS

The 6.3 l/sec (0.15 mgd) two-stage, packed-bed, downflow, granular activated carbon pilot plants were operated successfully for 33 months using unfiltered and unchlorinated activated sludge effluent from the Pomona Water Renovation Plant. Extensive operating data collected during the field study had demonstrated the stability of the two-stage carbon adsorption system in producing consistently effluents of excellent overall quality. The average effluent dissolved chemical oxygen demand (DCOD) and suspended solids concentration were 6.4 mg/l and 2.1 mg/l, respectively for the Filtrasorb 300 system and 6.2 mg/l and 2.0 mg/l, respectively for the Filtrasorb 400 system.

The overall organic removal through the Filtrasorb 300 system averaged 74.7% for total chemical oxygen demand (TCOD) and 73% for DCOD. The corresponding removal in the Filtrasorb 400 system was 78.6% for TCOD and 74.5% for DCOD. The DCOD removal efficiency through the two-stage systems has not changed significantly after several regeneration cycles. In the Filtrasorb 300 system, the DCOD removal decreased from 72.3% during the virgin cycle to 68.9% after four adsorption cycles. The corresponding values for the Filtrasorb 400 system were 72.5% in the virgin cycle and 71.7% after three adsorption cycles.

During the virgin cycle, the carbon capacity, expressed as kg DCOD removed/kg carbon, was 0.35 for the 300 system and 0.39 for the 400 system. About 25% decrease in carbon capacity occurred after four adsorption cycles in the 300 system, resulting in a fourth cycle capacity of 0.26. The carbon capacity of the 400 system decreased about 23% from an initial level of 0.39 to an apparent steady-state level of 0.30 after three adsorption cycles. The 400 system has approximately 13% more DCOD removal capacity than the 300 system. While the 400 system showed a slightly higher carbon capacity than the 300 system, the latter has the advantage of lower pressure loss and lower carbon loss during regeneration.

The carbon dosage in the 300 system increased from .038 kg/m^3 (320 lbs. carbon/million gallons) to .066 kg/m^3 (550 lbs. carbon/million gallons) after four adsorption cycles. In the 400 system, the carbon dosage increased from .034 kg/m^3 (280 lbs. carbon/million gallons) to .06 kg/m^3 (500 lbs. carbon/million gallons) at the end of the third adsorption cycle.

The first stage carbon columns, which served as deep bed filters and adsorbers, were routinely backwashed every two days with a volume of secondary effluent equivalent to 1.8% of the product water. The filtering action through the carbon column readily removed about 10 mg/l of the suspended solids from the secondary effluent feed, resulting in an average

net headloss buildup during the two-day backwash cycle of .26 kg/cm^2 (3.7 psig) for the 300 system and .4 kg/cm^2 (5.7 psig) in the 400 system. Of the TCOD removal in the two stage systems, 76.4% is removed in the first stage, primarily due to filtration.

An evaluation of the carbon characteristics following repeated thermal regenerations shows that after four adsorption cycles in the 300 system, the iodine number, methylene blue number and mean particle diameter decreased 29.5%, 27%, and 14.4%, respectively. The decrease in iodine number, methylene blue number and mean particle diameter after three adsorption cycles in the 400 system were respectively, 23%, 14.9%, and 0%. During the same period, the molasses number increased from 222 to 347 in the 300 system, and from 237 to 350 in the 400 system.

The ash content of the carbon increased with adsorption cycle from 5.5% to 12.9% in the 300 system and from 5.8% to 11.7% in the 400 system. The estimated average carbon loss during regeneration is 6.8% over four adsorption cycles in the 300 system and 7.6% after three adsorption cycles in the 400 system.

The estimated total treatment cost for a .44 m^3/sec (10 mgd),two-stage carbon adsorption system, designed to produce a produce water with TCOD of 12 mg/l and DCOD of 7-8 mg/l using an activated sludge plant effluent feed, is11.52¢/1000 gallons. The estimated cost is based on using the 300 system carbon (8 x 30 mesh) with a carbon dosage of .066 kg/m^3 (550 lbs. carbon/million gallons) and a carbon regeneration loss of 7% per cycle.

These costs are referenced to October 1973.

SECTION III

RECOMMENDATIONS

1. A similar study with single stage carbon contact should be conducted.

2. Evaluation of optimum backwash techniques should be conducted.

3. The effect of louding on carbon loss during regeneration should be evaluated.

SECTION IV

EXPERIMENTAL PROGRAM

PILOT PLANT DESCRIPTION

Carbon Contacting System

The carbon columns, comprising the two parallel two-stage adsorption systems were the same columns used in the earlier 4-stage carbon adsorption system evaluated in Pomona (3). The general layout of the carbon adsorption system is shown in Figure 1, together with the location of the carbon regeneration and air pollution control systems. Figure 2 shows the schematic flow diagram of the two-stage adsorption systems. The detail of a typical carbon column is illustrated in Figure 3. Each of the carbon columns was 1.83m (6 ft) diameter by 4.88m (16 ft) high and was designed for a working pressure of 3.52 kg/cm^2 (50 psig). Before the columns were reverted to the two-stage operational modes, the interior of columns II, III and IV were sandblasted and recoated with corrosion-inhibiting coating. Columns II and III were coated with bitumastic coal-tar epoxy and contactor IV with a polyester resin Ceilcote Flakeline 252. The two-stage system, consisting of columns II and III, contained 3164 kg (6960 lbs) of Calgon Filtrasorb 300 (8 x 30 mesh), while the other system with columns IV and V contained 3027 kg (6660 lbs) of Calgon Filtrasorb 400 (12 x 40 mesh).

The system containing 8 x 30 mesh carbon is designated as Filtrasorb 300 system, while the other column system with 12 x 40 mesh carbon is referred to as Filtrasorb 400 system. Table I shows the characteristics of the virgin carbon. The carbon bed was supported on an underdrain system consisting of two layers of stainless steel plate perforated with 0.79 mm (1/32") diameter holes and spot-welded together with the perforations off-set from each other.

The two-stage systems were operated in downflow mode at a constant rate of 6.3 l/sec (100 gpm) thereby providing a hydraulic loading of 2.38 l/sec/m^2 (3.5 gpm/ft^2) and an empty-bed contact time of 20 minutes per stage. The unchlorinated effluent from a 0.35 m^3/sec (8 mgd) activated sludge plant was pumped to the lead contactors without any pretreatment. The column feed entered the top of the contactors through an annular distribution ring containing twenty 2.54 cm (1 in) diameter holes around the circumference of each contactor. The wastewater flowed in series through each of the two contactors in the two-stage systems. The entrance annular ring was located about 1.07 m (3.5 ft) above the top of the carbon bed, thus providing for 35% bed expansion during backwashing. Each contactor was provided with a fixed surface wash mechanism mounted about 5.08 cm (2 in) above the unexpanded carbon bed to assist in the routine column backwashing.

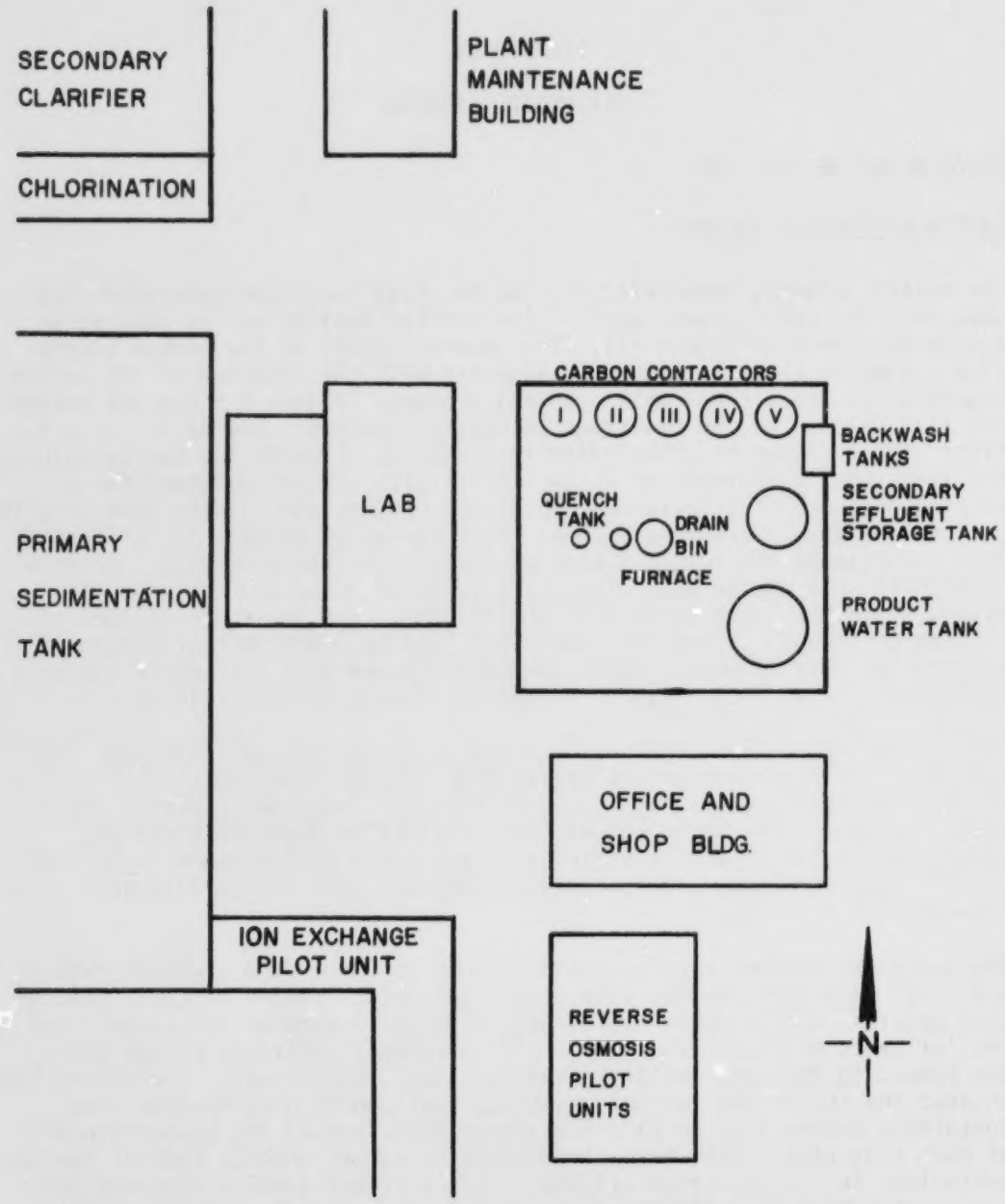

FIGURE 1: GENERAL LAYOUT OF THE CARBON ADSORPTION SYSTEM

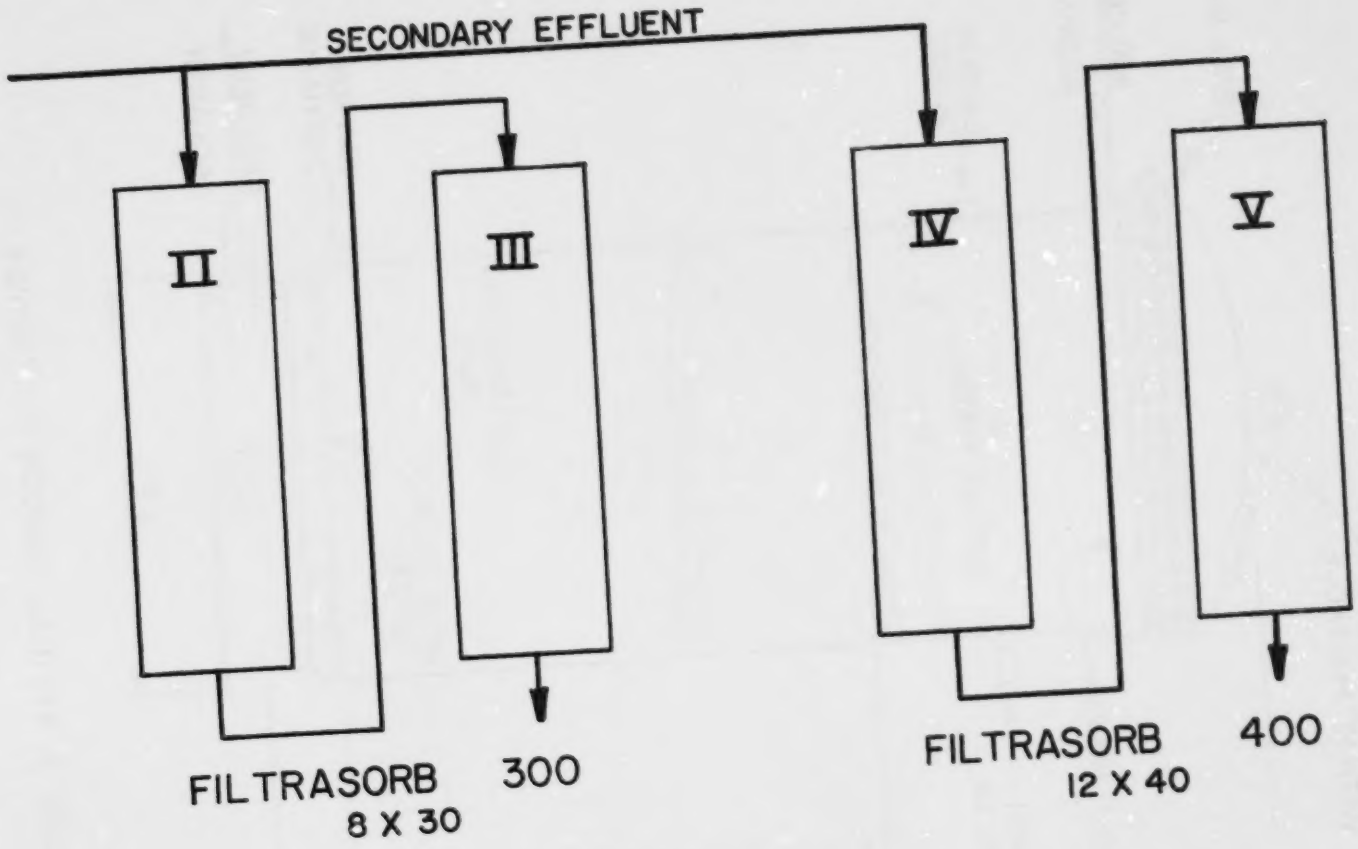

FIGURE 2: SCHEMATIC DIAGRAM OF THE TWO-STAGE ADSORPTION SYSTEMS

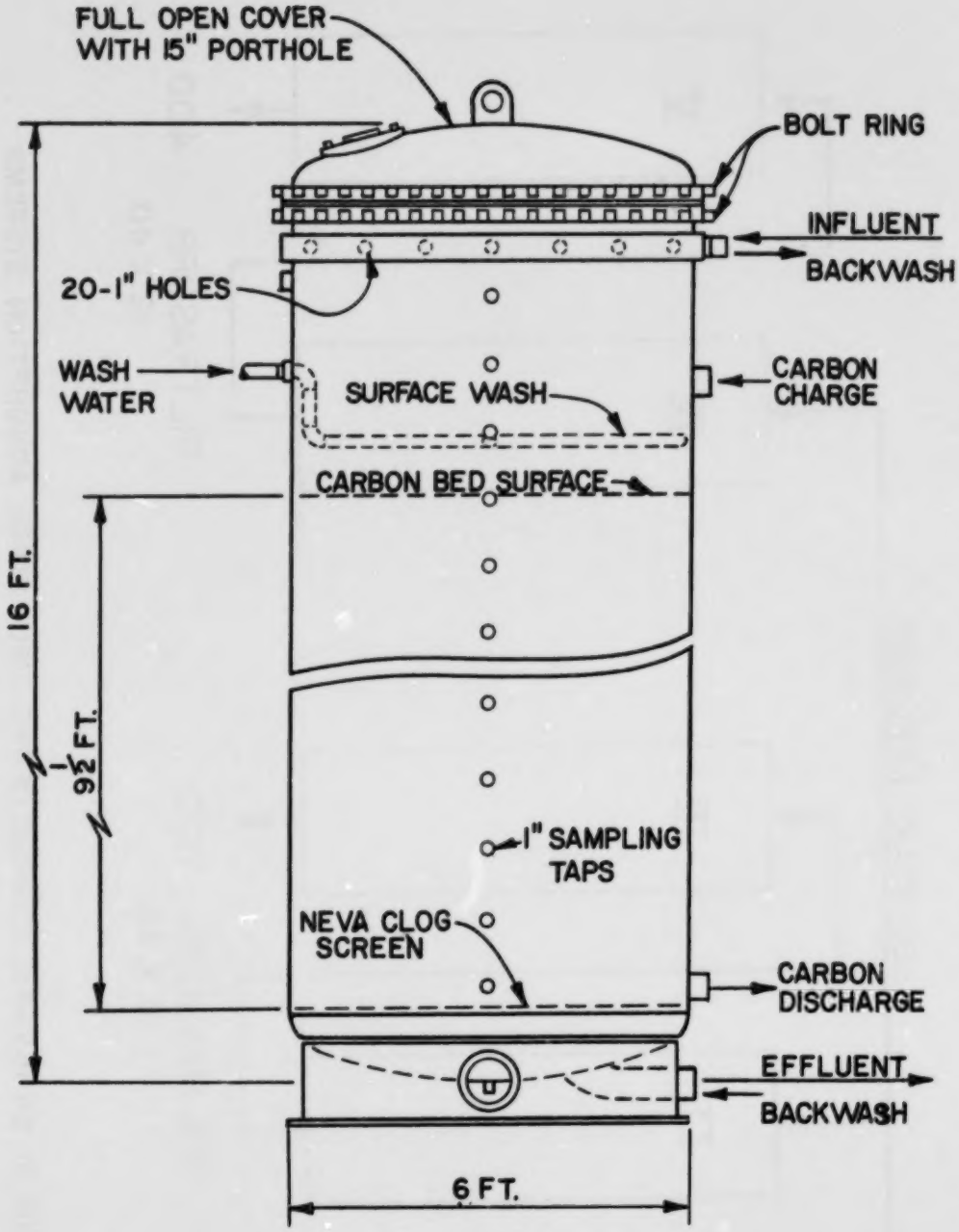

FIGURE 3: TYPICAL CARBON CONTACTOR

TABLE 1

VIRGIN CARBON CHARACTERISTICS

Carbon Characteristics		Calgon Filtrasorb 300 (8 x 30 mesh)	Calgon Filtrasorb 400 (12 x 40 mesh)
Iodine Number, mg/g		984	1062
Molasses Number		222	237
Methylene Blue No., mg/g		271	275
Apparent Density, g/cm^3		.484	.463
Ash, %		5.5	5.8
Mean Particle Dia., mm		1.6	1.0
Sieve Analysis: % Retained on No.	8	4.2	
	10	16.3	
	12	25.7	1.8
	14	18.7	10.3
	16	16.8	24.9
	18	10.2	20.8
	20	4.9	14.8
	30	2.6	20.6
	40		6.0
	Pan	0.6	0.8

Carbon Regeneration System

The regeneration furnace was a 76.2 cm (30 in) I.D., vertical, refractory-lined Bartlett-Snow-Pacific, Inc. multiple hearth furnace. As depicted in Figure 4, the furnace contained six hearths with two gas burners and steam inlets in each of the lower three hearths. The operation of the furnace was fully automatic, with push-button controls, safety equipment and a 12-pen furnace temperature recorder. The furnace was natural gas-fired with steam added to enhance the regeneration.

The partially dewatered, spent carbon from the drain bin was fed by a screw conveyor into the top of the regeneration furnace. The screw conveyor was provided with a variable speed drive so that the desired rate of carbon fed to the furnace can be accurately controlled. Regenerated carbon was discharged from the furnace through a 7.62 cm (3 in) diameter stainless steel chute leading from the bottom hearth into a quench tank, from which the regenerated, quenched carbon was continuously educted back to the original contactor.

Air Pollution Control System

In the course of the thermal regeneration of granular activated carbon spent on secondary effluents, severe air pollution problems could result. Experience with earlier field studies in Pomona (3) showed that the two major air pollutants associated with carbon regeneration were noxious odors and particulate emissions. In the early years of the pilot plant study several attempts were made to control these emissions through the use of a rotoclone and a direct-fired afterburner. The use of the rotoclone alone, or in series with the afterburner, was unable to meet the particulate discharge requirement of the Los Angeles County Air Pollution Control Districts (APCD). The operation of the afterburner alone at 927°C (1700°F) was very effective in odor removal, but only marginally effective in the control of particulate emissions. For four years until mid-1972 the carbon regeneration system operated with just the after-burner for air pollution control.

In view of the more stringent air pollution discharge requirements, it was decided in 1971 to install additional particulate emission control devices. The design and selection of the particulate control devices were made in cooperation with the APCD engineering staff.

The modified air pollution control system, shown schematically in Figure 5, consisted of a single cyclone dust separator, a baghouse, and a natural gas-fired afterburner. The new particulate emission devices were first placed in operation in July, 1972. During carbon regeneration, the flue gases from the furnace, together with particulates, first passed through the cyclone, which was designed to trap all burning particulates, 10 μ in diameter or larger before reaching the fabric filters in the baghouse.

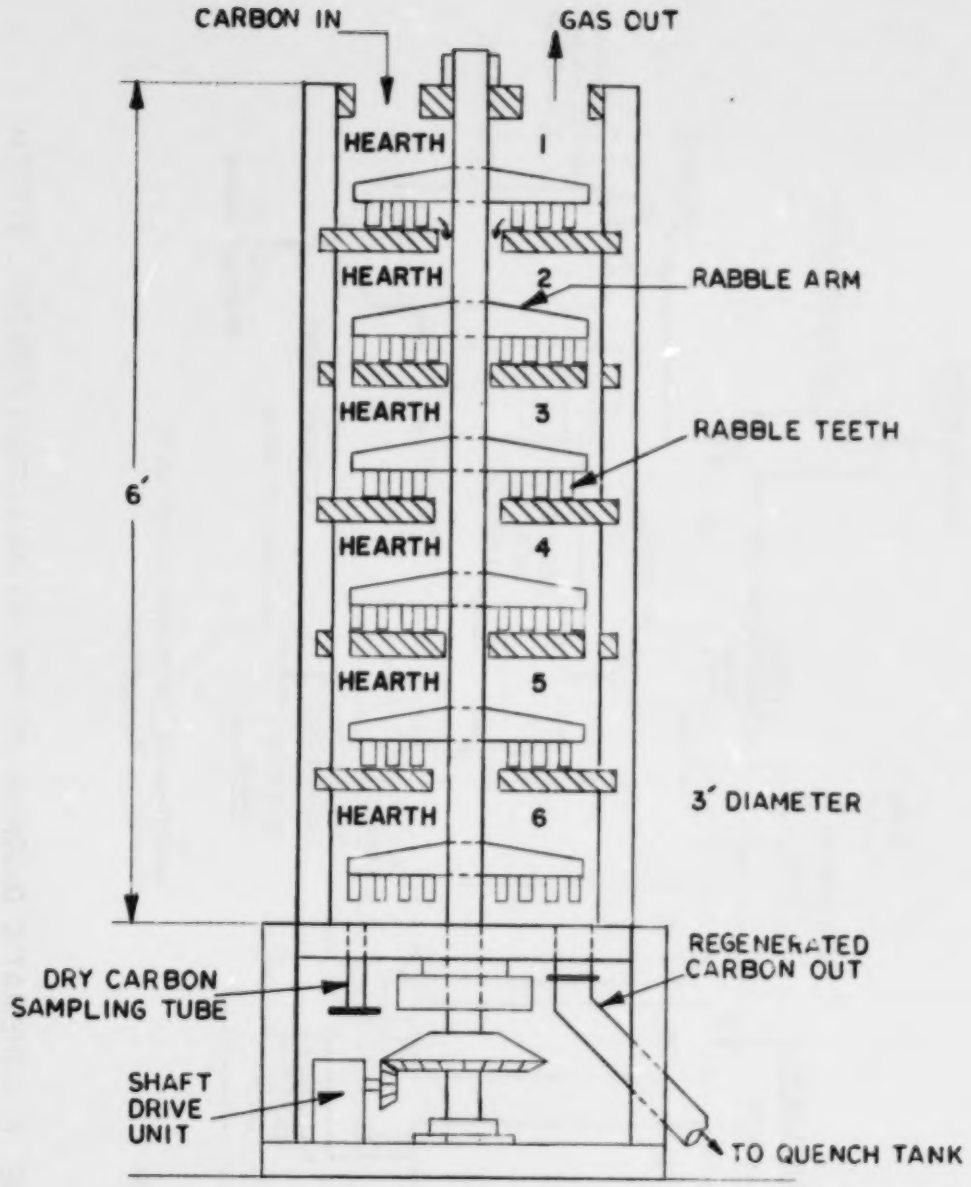

FIGURE 4: CROSS SECTIONAL VIEW OF THE MULTIPLE HEARTH FURNACE

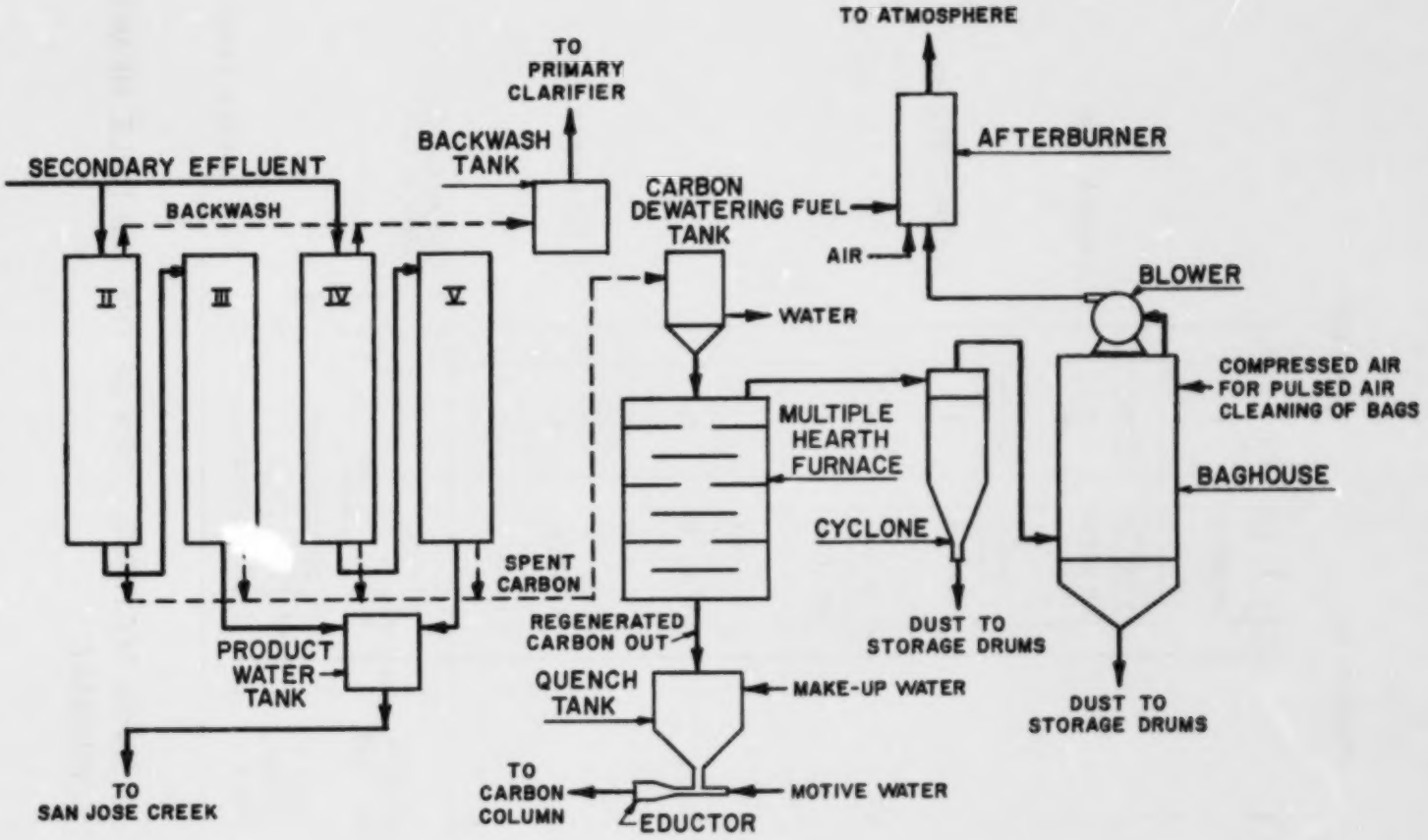

FIGURE 5. SCHEMATIC DIAGRAM OF THE AIR POLLUTION CONTROL SYSTEM

From the cyclone, the gases flowed through a series of ducts into a baghouse. The ductwork leading to the baghouse was provided with valved connection for dilution air addition. The baghouse was a reverse-air jet cleaned unit (Model 9-6-100 Mikro-Pulsaire Dust Collector manufactured by Mikropul Division of Slick Corporation), containing 9 Nomex felt fabric filter bags with a combined filter area of 5.95 sq m (64 sq ft). Each of the Nomex bags was 1.83 m (6 ft) long and was designed for an operating temperature of 204-218°C (400-425°F). The dust laden air entered the lower section of the baghouse and travelled upward through the fabric filter cylinder where the dust particles collected on the outside surface of the filter elements. A pull-through exhaust fan mounted on top of the baghouse provided the driving force for the dust flow through the system. As the dust mat builds up on the fabric surface, the pressure differential across the filter increased to a level where the deposited solids had to be removed by reverse air flow. In order to control the pressure drop through the filter within the desired limits of 2.54 cm (1 in) to 15.24 cm (6 in) water column, a cyclic timer periodically (2 to 45 second intervals) actuated the solenoid valves which delivered momentary surges of high pressure air 7kg /cm^2 (100 psig). The filtered gases from the baghouse then flowed through the afterburner which was operated between 704°C (1300°F) and 760°C (1400°F) for odor control, before final discharge into the atmosphere.

PILOT PLANT OPERATION

Column Designation

Each of the carbon columns in the two-stage adsorption systems was operated for extended periods in various operating sequences with carbon at different regeneration levels. In order to avoid confusion in discussing the column operation, it was necessary to identify the contactors according to the designation given in Figure 6. Thus, as an example, the operating sequence designated as IIOA, IIIOB would mean that the lead contactor or "A" position contactor is the Column II and the second stage or "B" position is Column III. Both columns contained virgin carbon which is represented by number "O". Since the columns were operated in a semi-counter-current mode, after the carbon in the lead contactor was regenerated, the column was placed on stream in the B position and the operating sequence would be designated as IIIOA, IIlB. Moreover, two adsorption sequences, that is, when each of the columns has been on stream in both "A" and "B" positions, represent an adsorption cycle.

Carbon Transfer and Regeneration

In liquid-phase adsorption system using beds of granular activated carbon, when the organic concentration of the effluent from the last stage exceeds a predetermined effluent limit, only the carbon layers in the inlet section are highly saturated. The carbon near the outlet end of the column system still has useable capacity. Thus, it is generally advantageous to operate the system in a counter-current manner in which only the highly saturated or exhausted carbon at the inlet section is removed from service for regeneration.

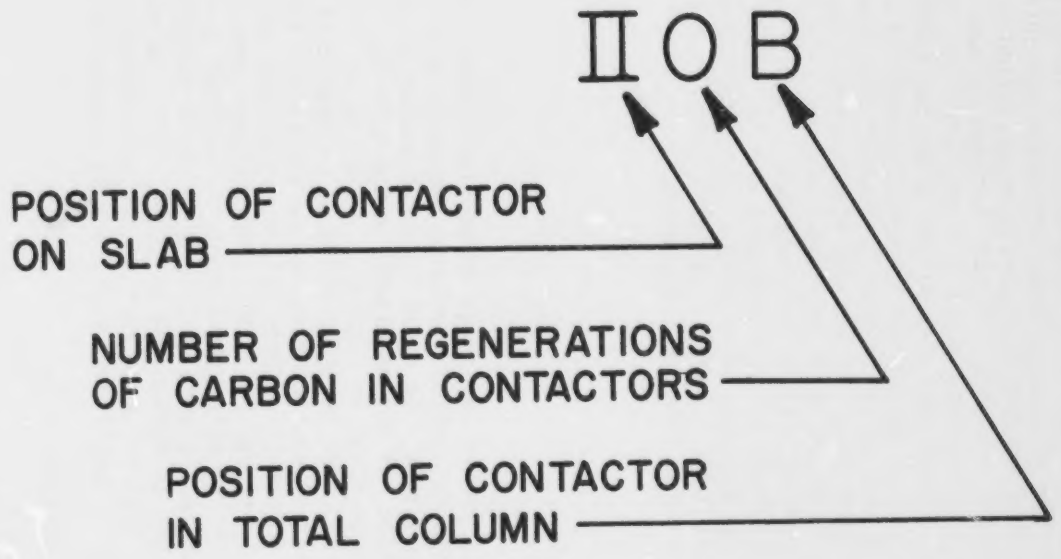

FIGURE 6: CONTACTOR IDENTIFICATION NUMBER

While there are a number of approaches to achieve a counter-current operation, the method adopted in the study was to operate two carbon columns in series. When the effluent from the two-stage system reached a TCOD of 12 mg/l, the lead carbon column was taken off stream in preparation for regeneration. The spent carbon was thoroughly backwashed before being hydraulically transferred as a slurry to an elevated dewatering bin. The backwash procedure was similar to that used for routine column backwash except for the fact that the last backwash step was prolonged to provide a total backwash water volume of about 79.5 m^3 (21,000 gallons). The dewatered spent carbon, with about 40% moisture content, was conveyed through a screw conveyor into a six-hearth furnace where it was regenerated at temperatures ranging from 899-982°C (1650-1800°F). As discussed previously, steam was added in the lower two hearths (hearths 5 and 6) in the amount of 0.6 kg steam/kg carbon to enhance the regeneration. The regenerated carbon was discharged from the furnace at the rate of 40.9-45.5 kg/hr (90-100 lbs/hr) into a quench tank from which it was continuously educted back into the original contactor. The regeneration of spent carbon in one contactor normally required about 72-95 hours to complete. After regeneration, the regenerated carbon was then backwashed with 56.8-75.7 m^3 (15,000-20,000 gallons) of secondary effluent to remove carbon fines and appropriate amount of make-up carbon added to replace the carbon lost during regeneration. The column, with the added make-up virgin carbon, was then briefly backwashed with 7.6-18.9 m^3 (2,000-5,000 gallons) of secondary effluent before the column was placed back in operation as a second stage in the system.

SAMPLING AND TESTING PROGRAM

Refrigerated 24-hr. composite samples of influent and effluent from each carbon column were collected automatically using timer-controlled solenoid valves. These samples were analyzed daily for turbidity and two to three times a week for TCOD, DCOD, suspended solids and color. About once a week, the samples were analyzed for ammonia, nitrite, nitrate, TOC, and methylene blue-active substances. Analytical determinations for pH and temperature were performed on grab samples three times a week.

The presence or absence of virus from the two-stage carbon effluent was determined once a week by analyzing a standard virus swab contacted with a stream of carbon effluent for a three-day period. In Figure 7 is shown a sketch of a flow-through virus sampling device. The average flow through the sampling module during the three-day contact period varied from .09 to .19 l/sec (1.5 to 3.0 gpm). All virus analyses were performed by the Los Angeles County Health Department.

All physical and chemical analyses were performed in accordance with the 12th edition of Standard Methods (4) or the FWPCA Methods for Chemical Analysis (5) unless otherwise specified. Turbidity was determined by the use of Rossum turbidimeter. TOC was measured by Beckman total organic carbon analyzer. The activated carbon analyses were performed using standardized procedure of the Pittsburg Activated Carbon Company.

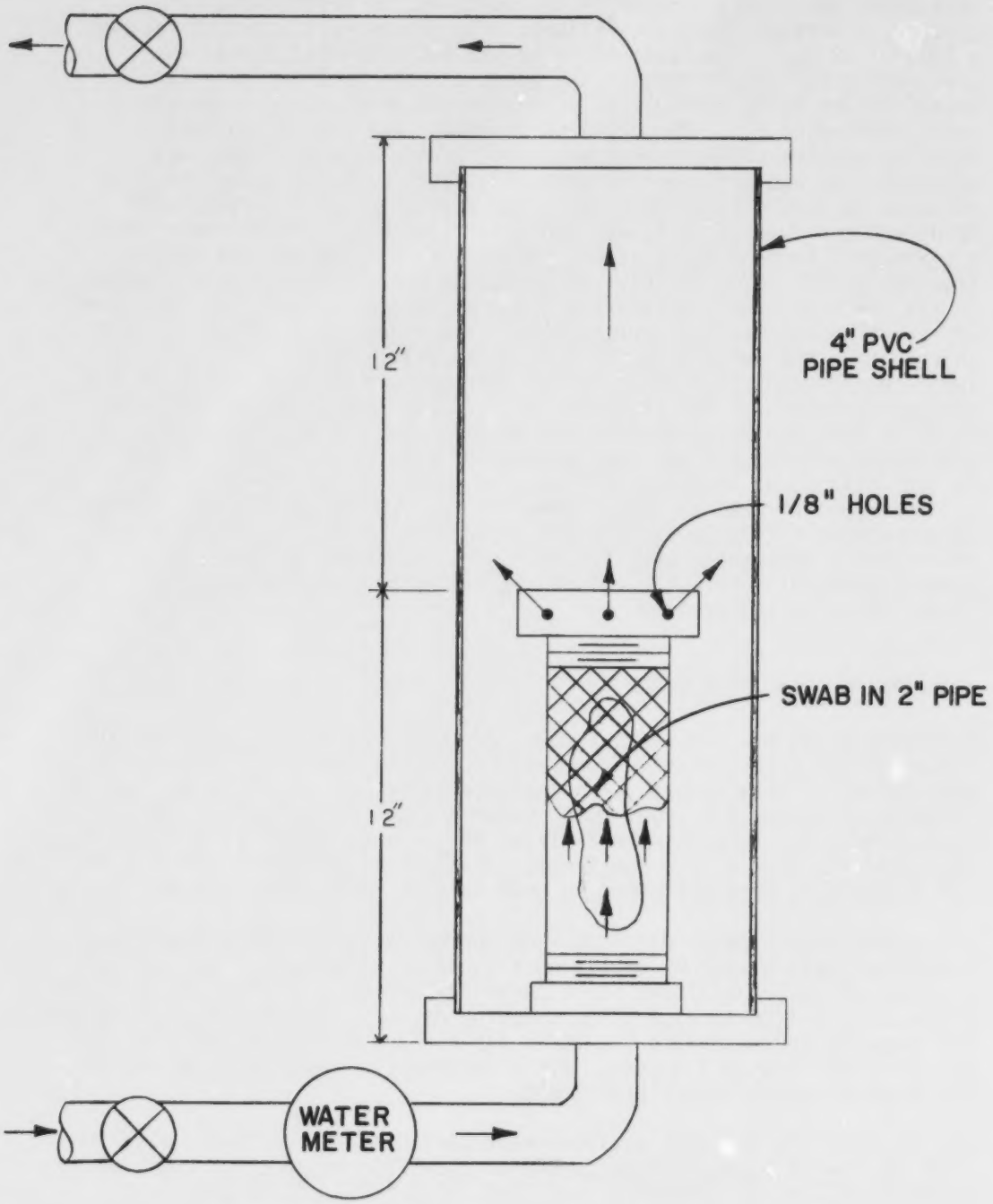

FIGURE 7: VIRUS SAMPLING CHAMBER

SECTION V

EXPERIMENTAL RESULTS

COLUMN BACKWASHING

Since the two-stage granular activated carbon columns were operated on packed-bed, downflow mode, the first stage carbon columns functioned as deep filter beds and as adsorbers. Thus, the application of unfiltered activated sludge effluent, with suspended solids concentration ranging from 10 to 20 mg/l, directly to the carbon columns led to progressive clogging of the beds with an attendant increase in headloss. To maintain proper column operation, the lead carbon columns were backwashed on a predetermined schedule with secondary effluent. During the course of the study, a number of backwash procedures as to frequency, type and duration, were tried. The actual backwash duration varied from 33 to 72 minutes depending on the procedure used. During the first three months of operation, the lead contactors ("A" position) were backwashed with about 56.8 m^3 (15,000 gallons) of secondary effluent once every 8 days or whenever the headloss reached 1.05 kg/cm^2 (15 psig). This infrequent backwashing operation led to heavy accumulations of biological solids in the carbon bed which were difficult to backwash. In addition, channeling in the carbon beds were observed. Consequently, the backwash frequency was increased to once every three days for about 1 1/2 months. Thereafter, the backwash frequency was increased to every 2 days, using a modified backwash procedure along with the use of a fixed surface spray mechanism. The backwash schedules for the original and the modified backwash procedure are shown in Figures 8 and 9. The modified backwash procedure, which had been used until the completion of the field study, had the following advantages over the original procedure:

1) The same degree of bed cleansing, as determined by visual observation of the backwash water and the headloss after backwash, was achieved with about half the volume of backwash water.
2) The system downtime during backwashing was reduced from 72 minutes or more to 33 minutes.
3) As indicated in Figure 9, the maximum backwash rate of 7.13 $l/sec/m^2$ (10.5 gpm/ft^2) was reached earlier, thus providing maximum bed expansion for a longer portion of the backwash operation.
4) The surface spray had been effective in breaking up the accumulated deposits on top of the carbon bed. In addition, the routine use of the surface spray during backwash helped in maintaining the carbon surface level and reduced channeling through the bed.

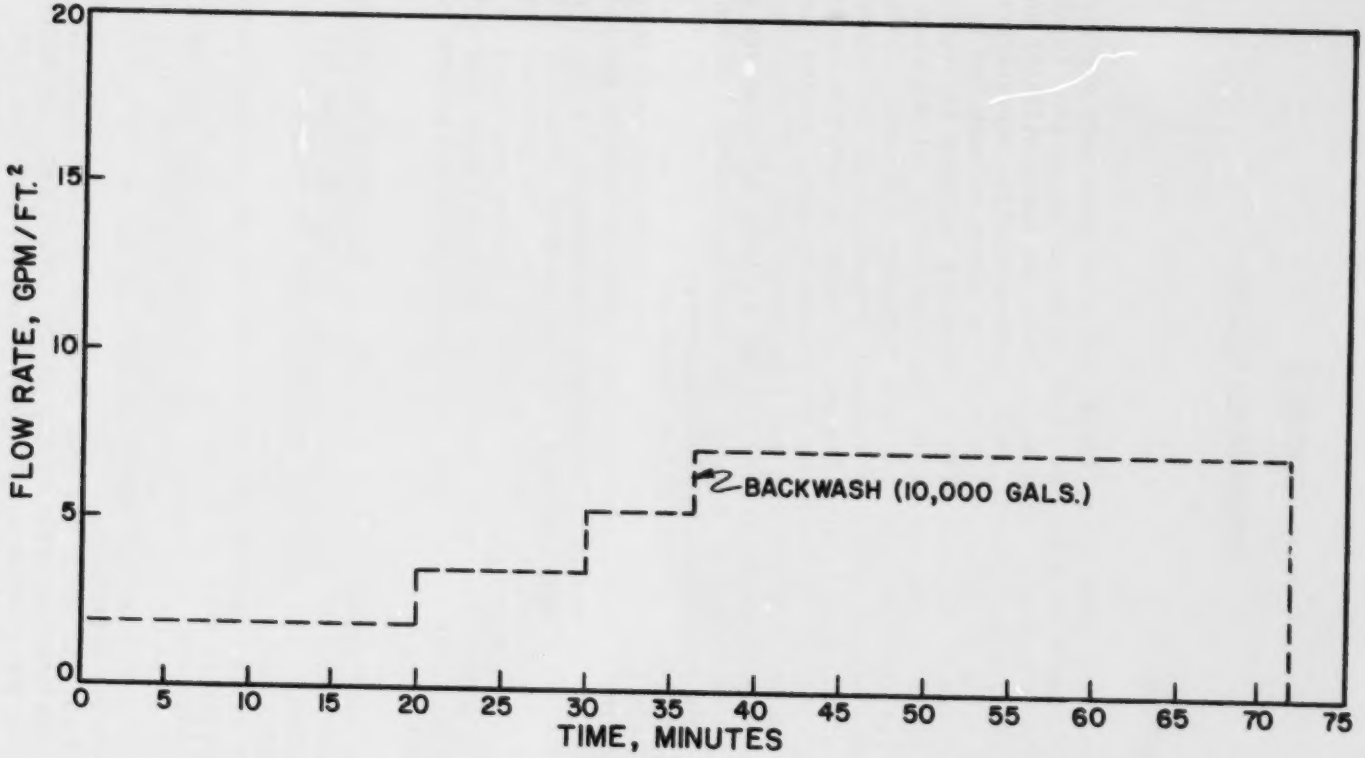

FIGURE 8: ORIGINAL BACKWASH SCHEDULE

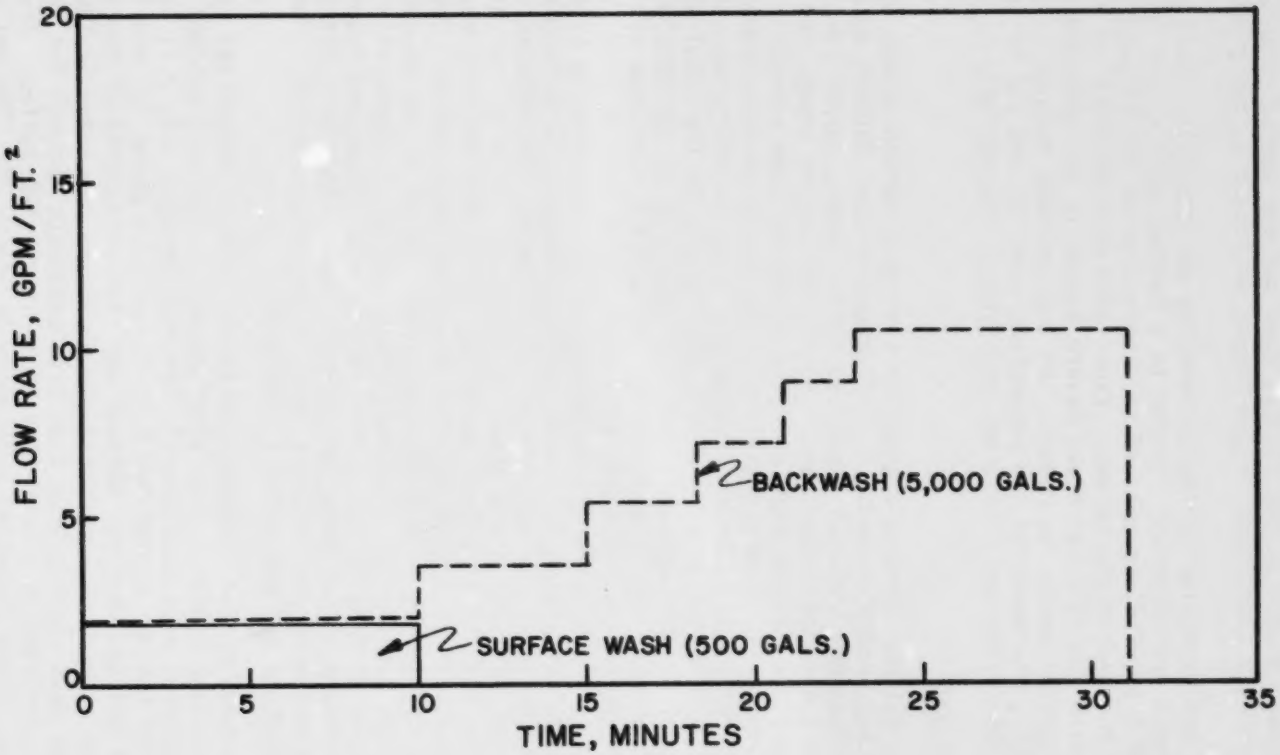

FIGURE 9: MODIFIED BACKWASH SCHEDULE

During backwashing, the backwash water was discharged into a holding tank designed to capture any accidental carbon spills and to allow visual observation of the clarity of the backwash water. The backwash water containing heavy accumulations of biological flocs and some carbon fines overflowed a weir in the holding tank and was pumped into the head end of the primary clarifiers of the Pomona activated sludge plant.

While a 35% bed expansion during backwashing was possible, routine column backwash operation was limited to a maximum upflow rate of 7.13 l/sec/m^2 (10.5 gpm/ft^2) because of the limitation on the structural strength of the underdrain screen. Experience with backwash at 9.5-10.86 l/sec/m^2 (14-16 gpm/ft^2) had shown severe damage to the underdrain screen which took several months to repair. At the backwash rate of 7.13 l/sec/m^2 (10.5 gpm/ft^2), the measured bed expansions were 10% and 15% in the Filtrasorb 300 system and Filtrasorb 400 system, respectively.

HEADLOSS BUILDUP

The headloss buildup through a granular activated carbon column operated on a packed-bed, downflow mode, is influenced by such factors as hydraulic surface loading, influent suspended solids level, carbon particle size and the length of filter runs. During the entire study, the two parallel two-stage systems were operated using the same secondary effluent feed under identical conditions of hydraulic surface loading, filter run lengths and backwash procedure. Thus, assuming that the contribution of biological growths within the columns on the total pressure drop were equal in both systems, then the magnitude of the pressure drop buildup through the lead columns during the two-day back-wash cycle would depend primarily on the media particle size. With this in mind, a regression analysis was performed to determine the relationship between column influent turbidity, T, and net pressure drop, P, in each of the two types of carbon at various regeneration levels. The results of the regression analysis, which are shown in Figure 10, clearly indicate that at a hydraulic loading of 2.38 l/sec/m^2 (3.5 gpm/ft^2) the 8x30 mesh carbon exhibited about 33% less headloss than the 12x40 mesh carbon. The figures further demonstrate the effect of regeneration level on the column headloss. For both systems the headloss, after the first regeneration, was higher compared to that in the virgin carbon. In subsequent regenerations, while the pressure losses were consistently lower, they were not significantly different from the pressure loss in the virgin carbon. The regression equations are presented in Table 2.

The effectiveness of the backwash operation could be determined by examining the column headloss just after backwash. In both systems the initial headlosses were consistently low and averaged only 0.11 kg/cm^2 (1.6 psig) for the 300 system and 0.13 kg/cm^2 (1.9 psig) for the 400 system. The average headloss through the lead contactors of the two-stage systems at various adsorption sequences are presented in Tables 3 and 4. The data clearly show that the headlosses through the 400 system were consistently higher than those in the 300 system. These data confirmed those presented previously in Figure 10. The net pressure losses through the lead contactors, which are attributed to the accumulation of suspended solids, averaged 0.26 kg/cm^2 (3.7 psig) for the 300 system and 0.40 kg/cm^2 (5.7 psig) for the 400 system.

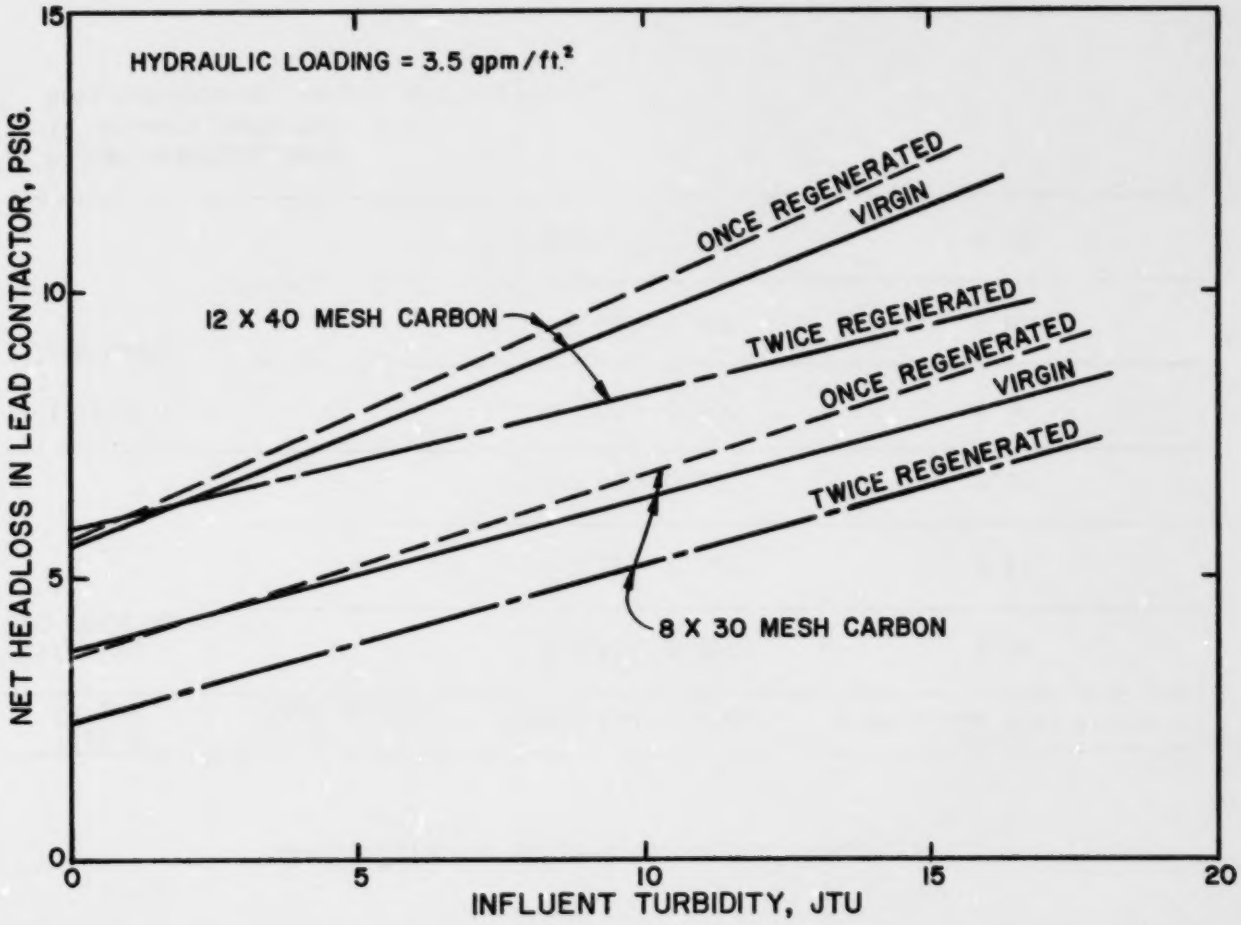

FIGURE 10: EFFECT OF INFLUENT TURBIDITY AND CARBON PARTICLE SIZE ON COLUMN HEADLOSS

TABLE 2

CORRELATION OF HEADLOSS WITH INFLUENT TURBIDITY

Type of Carbon	No. of Regenerations	Regression Equation	Correlation Coefficient
Filtrasorb 300 (8x30 mesh)	0	P=3.68 + 0.26T	0.44
	1	P=3.55 + 0.32T	0.54
	2	P=2.29 + 0.28T	0.66
Filtrasorb 400 (12x40 mesh)	0	P=5.52 + 0.40T	0.40
	1	P=5.60 + 0.45T	0.61
	2	P=5.71 + 0.25T	0.38

P= Net headloss, psig
T= Influent turbidity, JTU
Unit Conversion: psig x .070 = kg/sq cm

TABLE 3

HEADLOSS BUILDUP AND TURBIDITY REMOVAL IN THE
FILTRASORB 300 TWO-STAGE ADSORPTION SYSTEM

Run No.	Pressure Loss, psig				Turbidity, JTU			Total Volume Treated, million gallons
	1st Stage		2nd Stage		Influent	1st Stage Effluent	2nd Stage Effluent	
	BBW	ABW	BBW	ABW				
1	5.6	1.4	1.3	1.0	6.7	2.2	1.9	26.324
2	5.0	1.4	0.9	0.9	3.9	1.2	1.0	12.240
3	4.8	1.5	0.8	0.8	5.1	1.6	1.3	16.363
4	6.7	2.4	0.8	0.9	7.2	2.3	2.0	8.408
5	5.8	1.8	0.7	0.8	9.5	2.2	2.0	19.294
6	5.4	1.2	1.0	1.0	15.1	4.3	4.0	9.420
7	4.2	0.9	1.3	1.2	6.7	2.5	2.4	10.410
8	5.2	2.1	0.6	0.9	8.7	2.2	1.6	18.073

BBW = Before backwash
ABW = After backwash
Unit Conversions: psig x .070 = kg/sq cm
mil gal x 3785 = cu m

TABLE 4

HEADLOSS BUILDUP AND TURBIDITY REMOVAL IN THE
FILTRASORB 400 TWO-STAGE ADSORPTION SYSTEM

Run No.	Pressure Loss, psig				Turbidity, JTU			Total Volume Treated, million gallons
	1st Stage		2nd Stage		Influent	1st Stage Effluent	2nd Stage Effluent	
	BBW	ABW	BBW	ABW				
1	7.2	1.9	1.7	1.5	5.4	1.4	1.1	29.451
2	7.8	1.9	1.0	0.8	4.2	1.3	1.0	11.798
3	4.3	1.1	1.1	1.0	5.6	1.5	1.2	15.546
4	8.9	2.2	1.0	1.0	6.7	2.2	1.8	9.627
5	7.7	1.8	1.2	1.0	10.4	2.0	1.9	19.328
6	10.0	2.5	1.3	1.4	12.3	3.4	3.0	11.864

BBW = Before backwash
ABW = After backwash
Unit Conversions: psig x .070 = kg/sq cm
mil gal x 3785 = cu m

CARBON REGENERATION RESULTS

Control of the Regeneration Process

The economical use of granular activated carbon in wastewater treatment application demands that the exhausted carbon be repeatedly regenerated and then reused. The main goal of carbon regeneration is to effect maximum restoration of the exhausted carbon to its virgin properties. This goal is achieved by subjecting the carbon through a three-step process; namely: drying, baking and activating (6). The process variables, such as furnace temperature, carbon feed rate, and steam feed rate are closely controlled during regeneration in such a way as to effect maximum removal of adsorbed organics from the pores of the spent carbon, while at the same time, minimizing the damage to the basic pore structure of the carbon. In the activation step, the temperature is controlled automatically within 899-954°C (1650-1750°F), with steam added in hearths 5 and 6 to enhance regeneration.

In the course of regeneration, a number of control tests, consisting of the determination of apparent density, iodine number, methylene blue number and molasses number, were performed to monitor the quality of the regenerated carbon. The apparent density of virgin carbon normally ranged from 0.48 to 0.49 g/cm^3, which increased to about 0.59 when the carbon becomes exhausted. When the carbon is properly regenerated, the adsorbed organics are removed, thereby restoring the apparent density to the virgin level. The extent of recovery of the carbon adsorptive capacity was determined by running the decolorizing tests for iodine and molasses numbers.

The test for apparent density was determined routinely every hour whereas the test for iodine and molasses numbers were performed every 4 hours and 2 hours, respectively. The 6 to 8 grab samples of spent carbon, collected during carbon transfer, and the hourly samples of regenerated and quenched carbon were composited over the regeneration period and analyzed for apparent density, molasses number, iodine number, methylene blue number, ash content and mean particle size.

Effects of Regeneration

As a result of the various operations entailed in the thermal regeneration process, some loss invariably occur both in the carbon adsorptive capacity and also in the carbon quantity. Both these losses are of economic concern since they constitute a significant portion of the overall carbon regeneration cost. The carbon loss during each regeneration cycle has varied from 6.3% to 11% in the Filtrasorb 400 system and from 4.8% to 10.4% in the Filtrasorb 300 system. As used in this report, carbon loss is defined as the difference in the carbon volume in the contactor just before transfer to the drain bin and just after backwash of the regenerated carbon. The carbon loss, which is indicative of the carbon particle volume decrease, can be attributed either to the direct oxidation of the outer layers of the carbon granules

by the regenerating gases and/or to the normal handling attrition if the basic carbon structure has been weakened by internal overactivation. (6)

With the physical loss of carbon during regeneration, is the attendant decrease in the adsorptive capacity as measured by dissolved COD removal and iodine number. This observed loss in the carbon adsorptive capacity following repeated thermal regeneration is ascribed to the changes in the physical properties of the carbon. During regeneration, the complete removal of adsorbed organics from the carbon pores is really never attained. In addition to the ash buildup in the carbon pores, some carbon particles are unavoidably burned with the adsorbate thereby reducing further the total carbon surface area available for adsorption (6).

The changes in some of the physical properties of the carbon before and after several regenerations are shown in Tables 5 and 6. It is evident from the data that cyclic thermal regenerations has the effect of reducing the iodine number and increasing the molasses number and the % ash content. The increase in molasses number and the decrease in the iodine number is consistent with the fact that repeated thermal regenerations will cause a change in the pore size distribution of the carbon granule with the micropore structures being enlarged and thus producing a larger per cent of macropores. This shift in pore size distribution has the overall effect of reducing the total surface area of the carbon.

In this study, the iodine number was used as an index to measure the extent to which the carbon micropores were cleared of adsorbate during regeneration. The iodine number relates to the surface area of pores larger than 10A^{o} diameter. The iodine number of both spent and regenerated carbon are presented in Figure 11 for the 300 system and in Figure 12 for the 400 system. As indicated in Figure 11, the iodine number decreased 34% from a virgin level of 984 mg/g to 650 mg/g after four adsorption cycles in the 300 system. For the 400 system, the iodine number decreased from a virgin value of 1062 mg/g to 743 mg/g at the end of the third adsorption cycle. A continuing drop in the iodine number with adsorption cycle is apparent from the figures.

The ash content of the carbon, which measures the buildup of calcium and other inorganic residues, increased significantly from a virgin level of 5.5% to 12.5% after four cycles in the 300 system. For the 400 system, the corresponding ash buildup was from 5.8% to 11.4% after three cycles. This buildup of ash, along with the increase in the apparent density, following repeated thermal regenerations, is consistent with the decreasing trend in the iodine number. Figures 13 and 14 present the data on ash buildup and mean particle diameter as a function of the adsorption cycle. As evident from Figure 13, the mean particle diameter of the Filtrasorb 300 carbon decreased about 14% after four adsorption cycles. However, no further significant decrease in particle size occurred after the third cycle. In contrast, the mean particle diameter of the Filtrasorb 400 carbon remained practically unchanged through three adsorption cycles as shown in Figure 14.

In the course of thermal regeneration, some degree of internal overactivation occurs which results in an increase in macropore volume, that is, increase of pores larger than 30A^{o} diameter. The molasses number, which

TABLE 5

EFFECT OF REGENERATION ON THE PROPERTIES
OF FILTRASORB 300 CARBON
(8 x 30 mesh carbon)

No. of Regenerations		0	1	2	3	4
Adsorption Cycle			1	2	3	4
Iodine Number (mg/g)	(1)	668	652	567	484	
	(2)	984*	974	843	802	695
	(3)		911	817	760	651
Molasses Number	(1)	102	194	141	190	
	(2)	222*	265	333	374	347
	(3)		189	303	301	328
Methylene Blue Number (mg/g)	(1)	193	178	163	156	
	(2)	271*	271	228	231	198
	(3)		256	227	221	179
Apparent Density (g/cm^3)	(1)	.573	.575	.564	.585	
	(2)	.484*	.483	.493	.498	.508
	(3)		.495	.493	.499	.516
Ash, (%)	(1)	7	7.6	9.6	12.1	
	(2)	5.5*	8.1	10.3	10.9	12.9
	(3)		7.2	8.6	10.2	12.5
Mean Particle Diameter, (mm)	(1)	1.37	1.43	1.36	1.41	
	(2)	1.6*	1.37	1.46	1.31	1.37
	(3)		1.49	1.52	1.43	1.41

(1) Spent carbon
(2) Regenerated carbon before quenching
(3) Regenerated carbon after quenching
* Virgin carbon

TABLE 6

EFFECT OF REGENERATION ON THE PROPERTIES
OF FILTRASORB 400 CARBON
(12 x 40 mesh Carbon)

No. of Regenerations		0	1	2	3
Adsorption Cycle			1	2	3
Iodine Number (mg/g)	(1)	725	663	505	
	(2)	1062*	950	871	815
	(3)		926	823	743
Molasses Number	(1)	159	175	133	
	(2)	237*	286	285	350
	(3)		261	264	323
Methylene Blue Number (mg/g)	(1)	184	170	153	
	(2)	275*	238	234	234
	(3)		224	232	218
Apparent Density (g/cm^3)	(1)	.560	.555	.580	
	(2)	.463*	.479	.479	.496
	(3)		.472	.484	.504
Ash, (%)	(1)	6.8	8.2	10.1	
	(2)	5.8*	8.0	9.9	11.7
	(3)		7.8	9.2	11.4
Mean Particle Diameter, (mm)	(1)	1.04	0.99	1.01	
	(2)	1.0*	.99	1.02	1.0
	(3)		1.08	1.01	1.06

(1) Spent carbon
(2) Regenerated carbon before quenching
(3) Regenerated carbon after quenching

* Virgin carbon

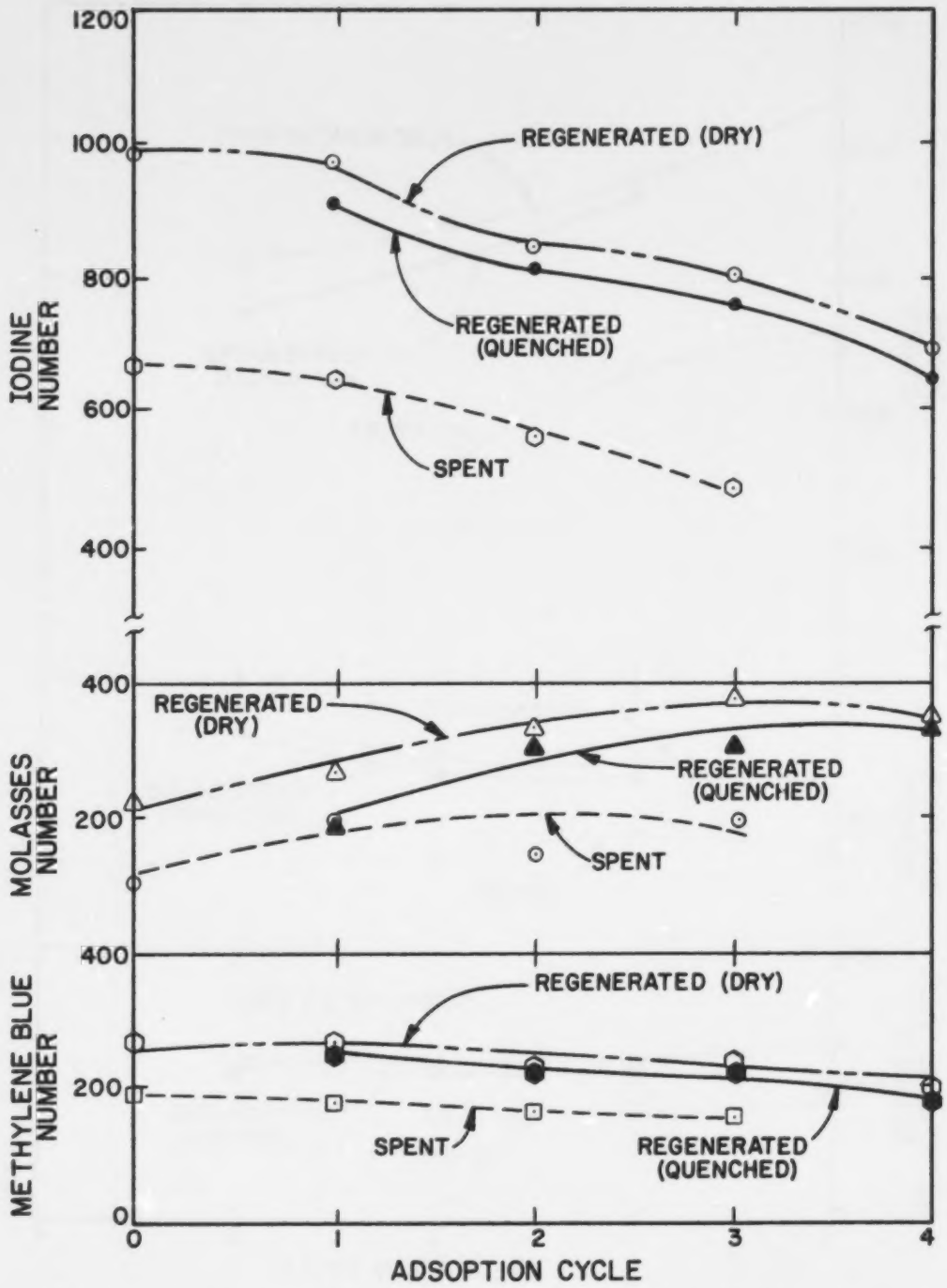

FIGURE II: EFFECT OF ADSORPTION CYCLE ON IODINE NO., MOLASSES NO., METHYLENE BLUE NO. OF FILTRASORB 300

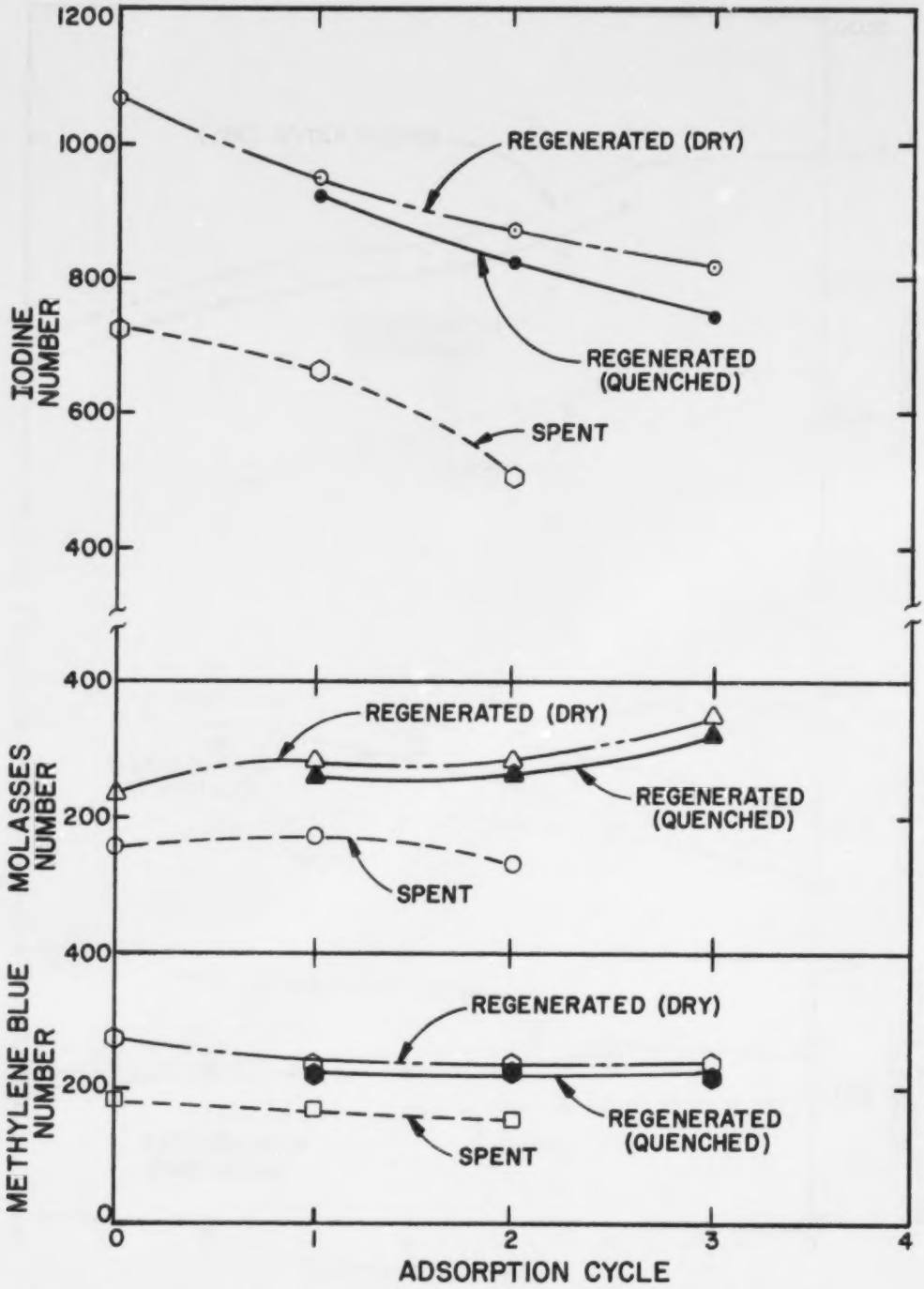

FIGURE 12: EFFECT OF ADSORPTION CYCLE ON IODINE NO., MOLASSES NO., METHYLENE BLUE NO., OF FILTRASORB 400

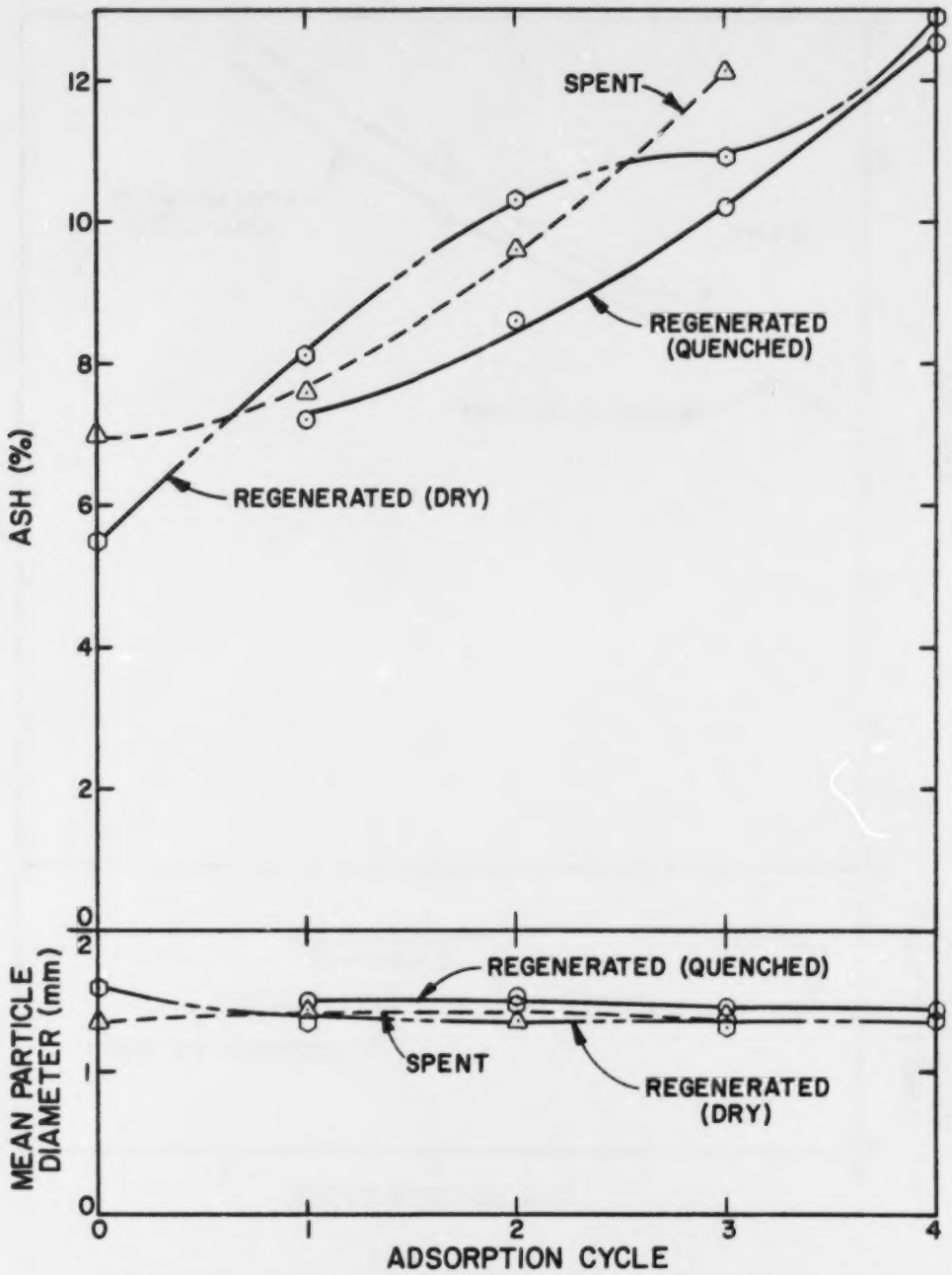

FIGURE 13: EFFECT OF ADSORPTION CYCLE ON ASH BUILDUP AND MEAN PARTICLE DIAMETER OF FILTRASORB 300

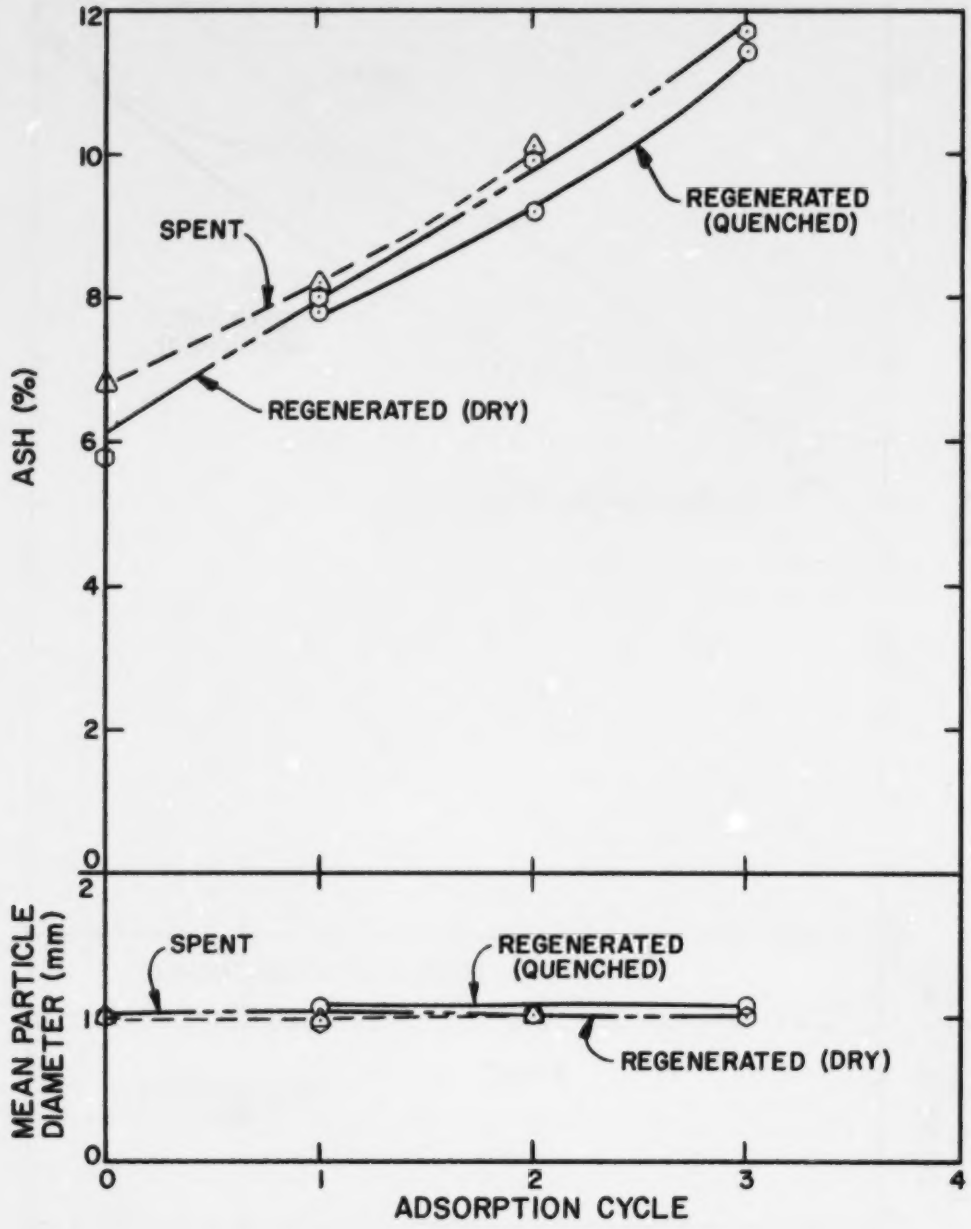

FIGURE 14: EFFECT OF ADSORPTION CYCLE ON ASH BUILDUP AND MEAN PARTICLE DIAMETER OF FILTRASORB 400

relates to the surface area of the pores larger than 28A° diameter, is taken as a measure of the pore enlargement. The molasses number of the virgin carbon has been measured in the range of 191 to 271 for Filtrasorb 400 and 190 to 238 for Filtrasorb 300. In all regeneration runs, the molasses number of regenerated carbon remained at high level, with a range of 231 to 323 for Filtrasorb 400 and 182 to 382 for Filtrasorb 300. Another parameter used to measure pore enlargement during regeneration, was the methylene blue number, which relates to surface area of carbon pores larger than 15A° diameter. In Figures 11 and 12 are shown the changes in the methylene blue, molasses and iodine numbers of carbon samples at various regeneration levels. The methylene blue and iodine numbers of regenerated carbon show a continuing decrease with regeneration level in contrast with the increasing trend of the molasses number. The relatively high molasses number and low iodine number indicate some internal damage to the carbon pore structure, resulting in a general shift to pores of larger size.

In examining the iodine number data presented in Figures 11 and 12, it is evident that a difference in iodine number of the regenerated carbon before and after quenching occurred consistently in all carbon regenerations. This observation has been confirmed by other investigators (6) who found reduction in iodine number of 25 to 69 units for virgin carbon Filtrasorb 400 heated to 954°C (1750°F) and quenched in water. Similarly, the molasses and methylene blue numbers of the regenerated carbon decreased after quenching.

In Tables 7 and 8 are summarized the operation data for the regeneration of each contactor in the two-stage adsorption systems. The data show some variations in the duration of regeneration, furnace loading, steam and fuel used. These variations in the furnace operation have been attributed primarily to the feeding conditions of the exhausted carbon.

As shown by the data in Table 7, the regeneration of spent carbon in column IIOA took an unusually long period to complete, due to operating problems with the furnace. Previous to this particular regeneration, the furnace had not been operated for about five months. In addition, there was a tremendous problem experienced in the transfer of carbon to the dewatering bin. Since the spent carbon in column II was not completely transferred to the drain bin, the original large quench tank was used as a temporary storage for the regenerated carbon. Thus, regeneration had to be stopped when the quench tank became full. The carbon remaining in the contactor was then manually transferred to the dewatering bin and, finally, regeneration was resumed. The regeneration of this last batch of carbon proceeded with great difficulty since it was "sticky" and difficult to feed. As anticipated, the fuel used in the regeneration was quite high and amounted to 10,000 BTU/lb. carbon. However, the carbon loss, which was estimated at 8%, was not unusually high compared with those observed in other regeneration periods.

In comparing the regeneration operating data in Tables 7 and 8 it is evident that invariably, the regeneration of Filtrasorb 400 carbon proceeded at a relatively shorter time with less fuel consumption than that of Filtrasorb 300. This difference in operating data was attributed to

TABLE 7

FURNACE OPERATING CONDITIONS DURING REGENERATION
OF 300 SYSTEM CARBON

Column Designation	Before Regeneration / After Regeneration	II0A / II1B	III0A / III1B	II1A / II2B	III1A / III2B	II2A / III3B	III2A	II3A / II4B	III3A / III4B	Average
Regeneration time (hours)		121	91	99	87	95	82	80	96	93.8
Furnace Loading (lbs. carbon/hr.)		57.5	76.5	71	89	81.6	97	99	82	81.7
Steam Used (lbs./lb. carbon)		.77	.69	.74	.59	.64	.54	.52	.64	.64
Fuel Used (BTU/lb. carbon)		10,000	6350	6160	5206	5080	4125	5219	6503	6080
Average Hearth Temp. (°F)										
Hearth No. 1		640	640	760	705	691	644	630	602	664
2		845	840	935	870	857	810	802	827	848
3		1090	880	1160	1049	1039	973	964	1015	1021
4		1635	1425	1665	1618	1631	1575	1514	1616	1585
5		1660	1500	1755	1758	1733	1713	1681	1731	1691
6		1660	1625	1730	1776	1689	1740	1708	1753	1710
Afterburner		1700	1750	1770	1782	1788	1764	1277	1440	1659
Carbon Loss, %		8	8	4.8	10.4	7.3	6.5	6.9	7	6.8

Unit Conversions: lb x .454 = kg
BTU/lb x .556 = cal/g
(°F-32) .555 = °C

TABLE 8

FURNACE OPERATING CONDITIONS DURING REGENERATION
OF 400 SYSTEM CARBON

Column Designation: Before Regeneration / After Regeneration	IV0A / IV1B	V0A / V1B	IV1A / IV2B	V1A / V2B	IV2A / IV3B	Average
Regeneration Time (hours)	72	78	82	97	78	81.4
Furnace loading (lbs. carbon/hr)	92.5	85.5	81	71	93	84.6
Steam used (lbs./lb. carbon)	.49	.62	.65	.74	.57	.61
Fuel used (BTU/lb. carbon)	5500	5350	5690	4660	4576	5155
Average Hearth Temp. (°F)						
Hearth No. 1	620	630	650	635	641	635
2	860	800	830	800	808	820
3	1070	1050	1000	955	966	1008
4	1725	1700	1580	1540	1571	1623
5	1840	1760	1730	1645	1724	1740
6	1770	1780	1720	1595	1726	1718
Afterburner	1820	1850	1775	1750	1796	1798
Carbon Loss, %	11	6	8.4	6.3	6.6	7.6

Unit Conversions: lb x 0.454 = kg
BTU/lb x .556 = cal/g
(°F-32) x 0.555 = °C

the fact that the regeneration of the 400 carbon always followed the regeneration of the 300 carbon. Thus, whatever furnace operating difficulties encountered during the regeneration of the 300 carbon were corrected before the regeneration of the 400 carbon.

As indicated in Table 7, before the baghouse was installed for air pollution control, the afterburner had been maintained at a temperature ranging from 927°C (1700°F) to 1010°C (1850°F). During the last two regenerations of the 300 carbon the baghouse system was used in series with an afterburner. In this operating mode, the afterburner temperature was reduced to an average of 738°C (1360°F), instead of the previous average level of 971°C (1780°F), since the afterburner was then used primarily for odor control.

Air Pollution Control Emission Data

The gases discharged from the top hearth of the regeneration furnace contained both fine carbon particulates and obnoxious smelling substances. These air pollutants were controlled through an air pollution control system consisting of a baghouse for particulate removal and an afterburner, operated in series with the baghouse, for odor control. The use of this air pollution control system, which was placed in operation for the first time during the regeneration of column II 3A, has made it possible to fully comply with particulate emission and other discharge requirements of the Los Angeles County Air Pollution Control District.

During the first 48 hours of operation of the baghouse, the filter ratio ranged from 12.7 to 13.7 l/sec/m^2 fabric area (2.5 to 2.7 cfm gas/ft^2) and averaged 13 l/sec/m^2 (2.55 cfm/ft^2). The corresponding average pressure drop through the baghouse was 4.06 cm (1.6 in) of water column (W.C.). In the remaining 32 hours of the regeneration, the pressure drop through the baghouse increased markedly with a range of 9.14 to 13.7 cm (3.6 to 5.4 in). As expected, the filter ratio decreased correspondingly to an average of 10.2 l/sec/m^2 (2 cfm/ft^2 fabric). It should be pointed out that at the start of the baghouse operation when the pressure drop through the filters was low, the reverse air jet cleaning cycle was made longer. The reverse jet was not operated during the first 6 hours in order to build up sufficient dust mat on the fabric filters which would improve the particulate removal efficiency. As the pressure drop increased to 7.6 cm (3 in), the cyclic timer was activated to pulse initially every 45 seconds. Thereafter, as the resistance increased, the cleaning cycle was increased to every 2 seconds intervals.

In the first day of operation of the air pollution control system, it became apparent that condensation was a serious problem in the baghouse, especially in the evening when the baghouse temperature ranged from 46.1 to 48.9°C (115 to 120°F). Because of the moist environment within the baghouse, blinding of the fabric filters occurred as indicated by the high headloss of 13.7 cm (5.4 in) through baghouse. The condensation problem was corrected by partially insulating the ductworks and the baghouse, thus maintaining about 93.3°C (200°F) in the baghouse during the evening.

Maintaining a high temperature in the baghouse, as long as it is below the fabric filter bags operating temperature of 218.3°C (425°F) is desirable for two reasons. Firstly, as previously mentioned, condensation with its attendant problems could be obviated. Secondly, since the filtered gases from the baghouse are subsequently conveyed to the afterburner for odor control, less amount of fuel would be required in operating the afterburner at 704.4 to 760°C (1300-1400°F). In subsequent regenerations, the ductworks and the baghouse were completely insulated with fiberglass, thus making it possible to maintain baghouse temperature range of 148.9 to 162.8°C (300 to 325°F). While it is advantageous to have a high temperature in the baghouse, due precaution must be exercised to prevent the temperature from rising to within 10 to 37.8°C (50-100°F) of the critical fabric filter design temperature. Thus, to minimize the danger of burning the filter bags, the baghouse inlet was equipped with a valved side connection for dilution air addition. Under normal furnace operating conditions, the dilution air inlet valve was maintained closed. However, when the baghouse temperature increased beyond 162.8°C (325°F), which could be caused by a disruption of the carbon feed rate to the furnace, the dilution air valve was manually opened for such a duration as required to restore the baghouse temperature within 148.9 to 162.8°C (300-325°F).

The performance of the various components of the air pollution control system was evaluated during two regenerations by test engineers from a local testing laboratory. During the first field evaluation of the air pollution control system, an engineer from the Los Angeles County APCD was also present to ascertain that correct sampling and testing procedures were followed. In evaluating the system, the gas at the inlet to the baghouse and at the outlet of the afterburner were tested for gas flow rate, temperature, particulate matter, volatile hydrocarbons, oxygen, carbon dioxide, carbon monoxide, water vapor, oxides of nitrogen, oxides of sulfur and odor number. The gas from the baghouse outlet was subjected to the same tests as mentioned above, except for oxides of nitrogen and odor number.

The gas flow measurements were made with a standard pilot tube and an inclined manometer. Temperatures were measured with a chromelalumel thermocouple and a portable potentiometer. Gas velocity and temperature traverse measurements were made at the afterburner outlet only. The particulate matter was collected using wet impingement in Smith-Greenberg impingers followed by Gelman filter holders equipped with glass fiber backup filters. Glass probes provided with glass ball joints were used as sampling probes. Samples for the determination of volatile hydrocarbons and fixed gases were taken in evacuated stainless steel tanks. Sulfur dioxide was collected, using a filter, followed by an impingement train containing hydrogen peroxide. Oxides of nitrogen were analyzed, using the phenol-disulfonic acid method. Orsat analyses were performed on the hydrocarbon samples and on the nitrogen oxide flasks.

A summary of the emission data from the various components of the air pollution control system is presented in Table 9. During the regeneration of Column II3A, particulate emission, as well as odor from the afterburner, were very low and averaged only .041 kg/hr (0.09 lb/hr) and

TABLE 9

SUMMARY OF AIR POLLUTION CONTROL SYSTEM PERFORMANCE

Parameters	APCD Emission Limit	Regeneration of Column II3A			Regeneration of Column III3A		
		Baghouse		After-Burner Outlet	Baghouse		After-Burner Outlet
		Inlet	Outlet		Inlet	Outlet	
1. Particulate Matter							
Concentration, grains/SCF	0.20	0.987	0.298	0.046	1.82	0.47	0.075
Emission rate, lbs/hr	1.00	0.890	0.266	0.09	2.17	0.80	0.24
2. Oxides of Nitrogen, (NO_x)							
Concentration, ppm dry	225	94		166	40	--	180
Emission rate, lbs/hr				0.48	0.028	--	0.40
3. Oxides of sulfur (SO_2)							
Concentration, ppm SO_2	0.2%			217	nil	nil	149
Emission rate, lbs/hr				0.88	--	--	0.57
4. Hydrocarbons							
Concentration, ppm C		3900		660	740	561	nil
Emission rate, lbs/hr C		0.88		0.50	0.20	0.21	--
5. Carbon Monoxide (CO)							
Concentration, % Volumedry		0.56		0.20	1.36	.86	0.11
6. Odor							
Odor units/SCF		20,000		10	20,000		20
7. Gas Flow							
Temperature, °F		345	140	1000	352	159	1148
SCFM		104	104	392	139	198	376

Unit conversions: lb/hr x 0.454 = kg/hr
(F°-32) x .555 = °C

10 odor units/scf, respectively. Based on particulate emission rate data, the baghouse removed only 70% of the dust, which was significantly below the design capability of about 99% removal. The total actual weight of dust collected from the baghouse over the 80 hour period of regeneration was 26.8 kg (59 lbs), which was about 20% more than the removal estimated from the emission rate data. This discrepancy, however, is not unreasonable, considering that the emission data represented samples collected over 45 to 60 minutes sampling period. In addition, difficulties were encountered in collecting representative samples from the baghouse inlet due to excessive condensation in the duct. The dust collected from the cyclone separator amounted to 6.36 kg (14 lbs).

In Table 9 are also shown the results of the second evaluation of the air pollution control system during the regeneration of column III3A. The data indicated that all emission parameters, such as particulate matter, oxides of nitrogen, oxides of sulfur, hydrocarbon and odor number, were all in full compliance with the local air pollution control requirements.

SYSTEM PERFORMANCE

Organic Removal

The COD test, both total and dissolved, was used in this study as the primary parameter in evaluating the performance of the activated carbon pilot plant in the removal of organic materials from the activated sludge plant effluent. The total COD (TCOD) removal patterns in each column for all adsorption sequences are presented in Figures 15 through 18 for the Filtrasorb 300 and Figures 19 through 20 for the Filtrasorb 400 system. In these figures, the top curves represent the column influent (secondary effluent) quality and the bottom curves designated with letter "B" represent the two-stage system effluent. The bands between the curves represent the amount of TCOD removed in each carbon column. The vertical bands between each adsorption run or sequence designate the time when the first stage or "A" position contactor was taken off-stream for regeneration. In examining these figures, it is evident that the major portion of the TCOD was removed in the first stage contactor as reflected by the width of the top bands compared with the bands between curves A and B. This is expected, since the first stage carbon column acted as an efficient filter capable of removing substantial amounts of the influent suspended material.

In determining the amount of COD applied and removed by each contactor, the areas bounded by the curves were determined using a planimeter. Although not presented in this report, DCOD removal patterns similar to those for TCOD, were also plotted from which the DCOD loading and removal capacities were calculated from the planimetered area.

The effect of repeated thermal regeneration or adsorption cycles on the DCOD removal capacity of the two-stage carbon adsorption systems are presented in Figure 21. For the Filtrasorb 300 system, the carbon capacity, expressed in kg DCOD removed/kg carbon, has decreased about 25% after

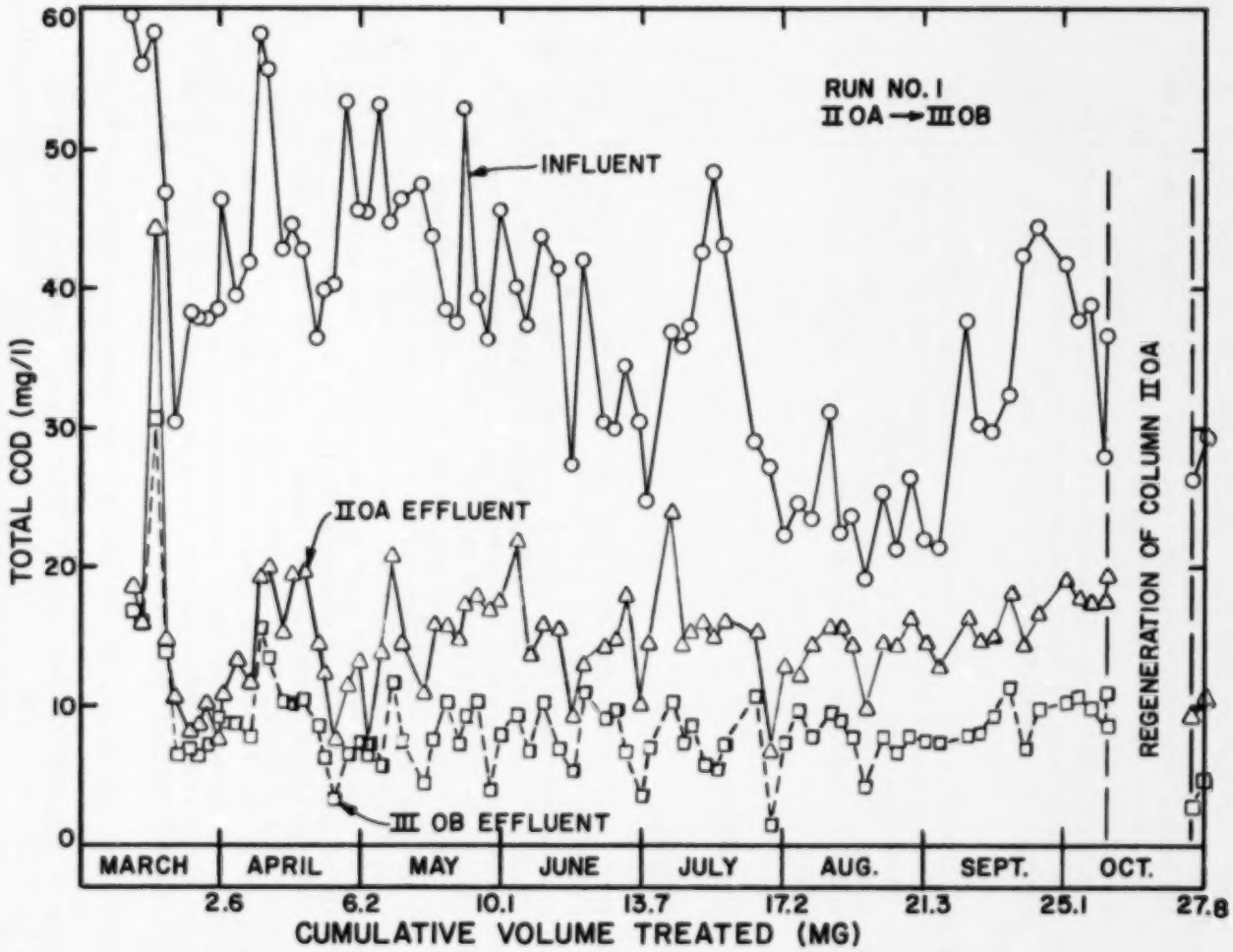

FIGURE 15: COD REMOVAL IN THE FILTRASORB 300 SYSTEM (Run 1)

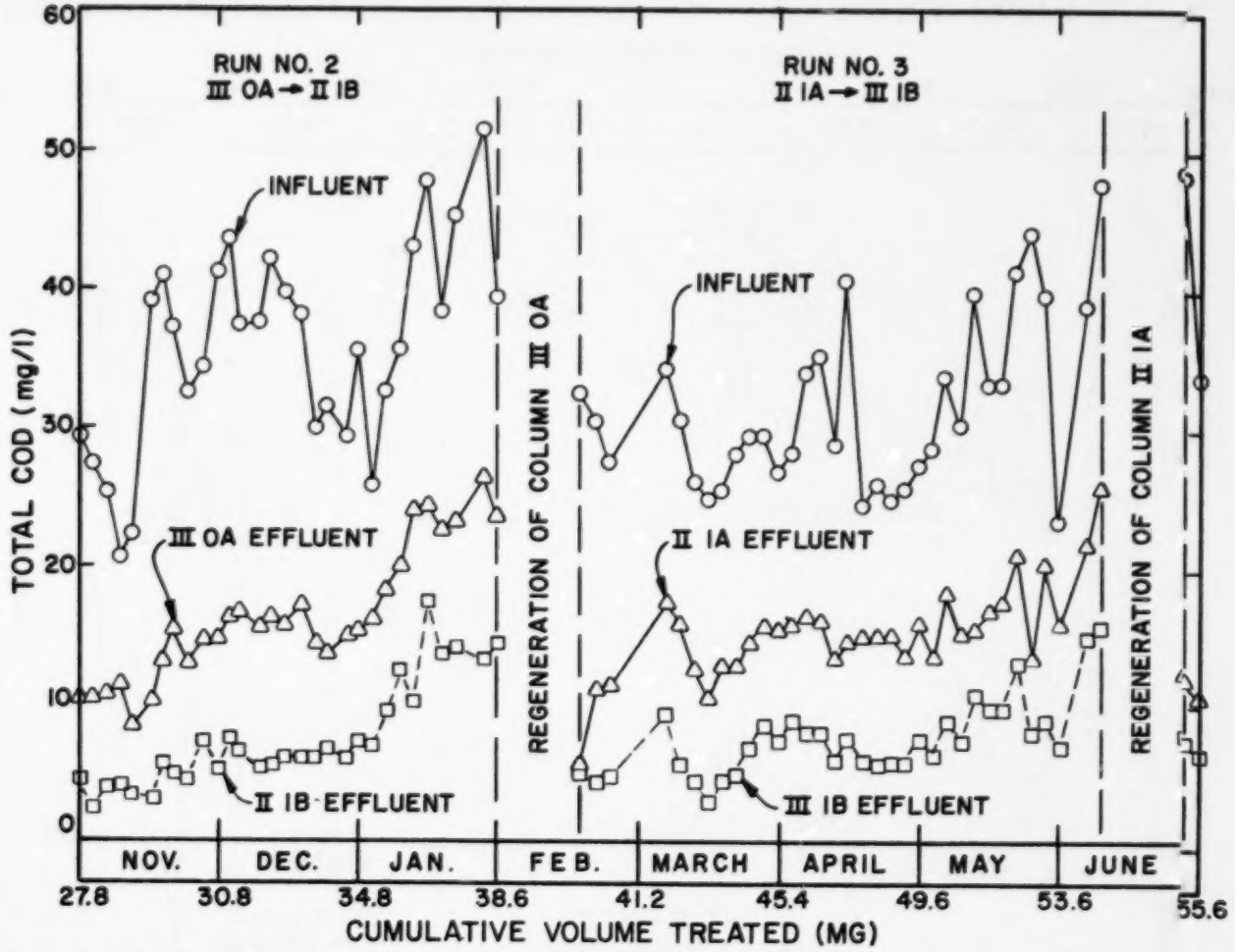

FIGURE 16: COD REMOVAL IN THE FILTRASORB 300 SYSTEM (Run 2-3)

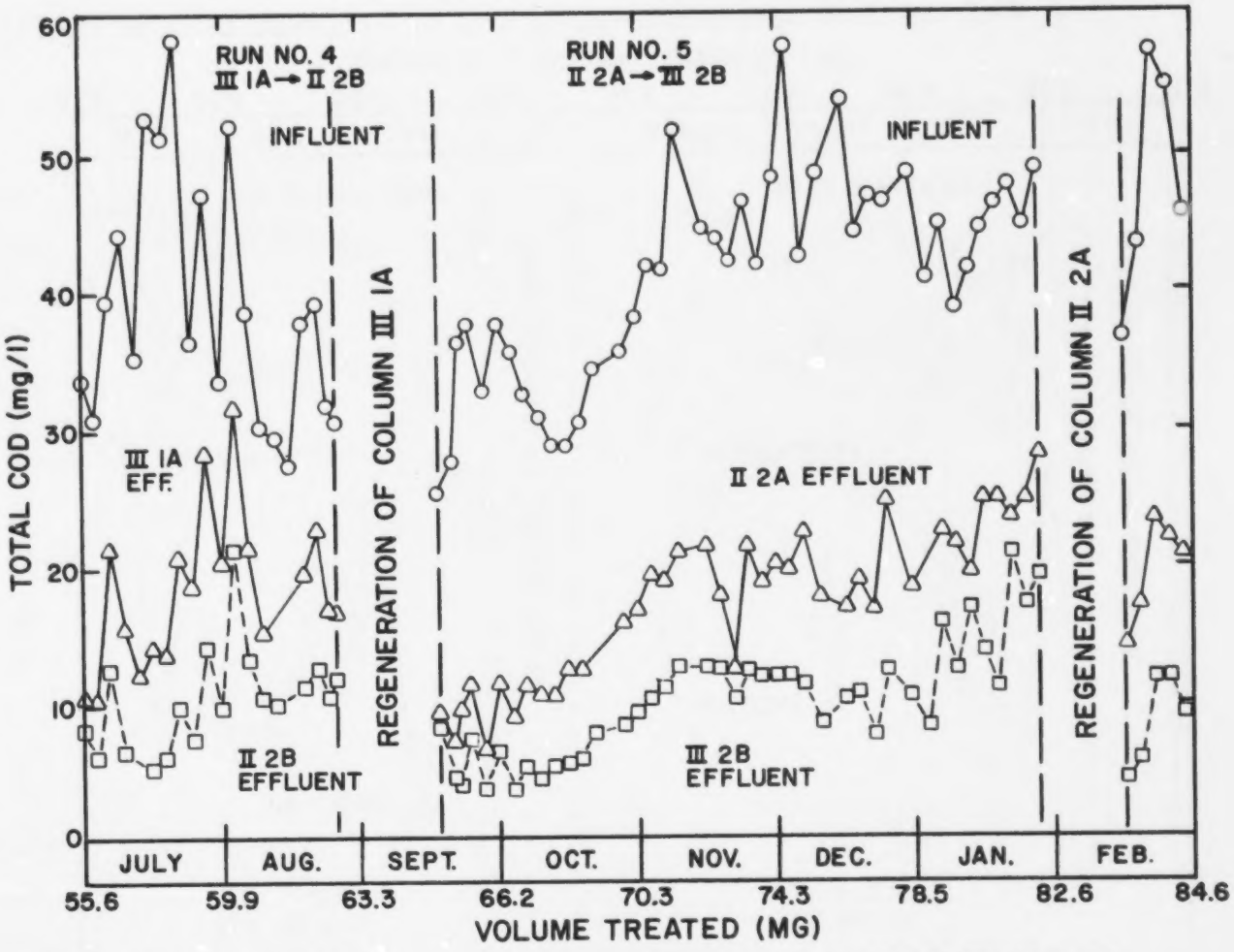

FIGURE 17: COD REMOVAL IN THE FILTRASORB 300 SYSTEM (RUN 4-5)

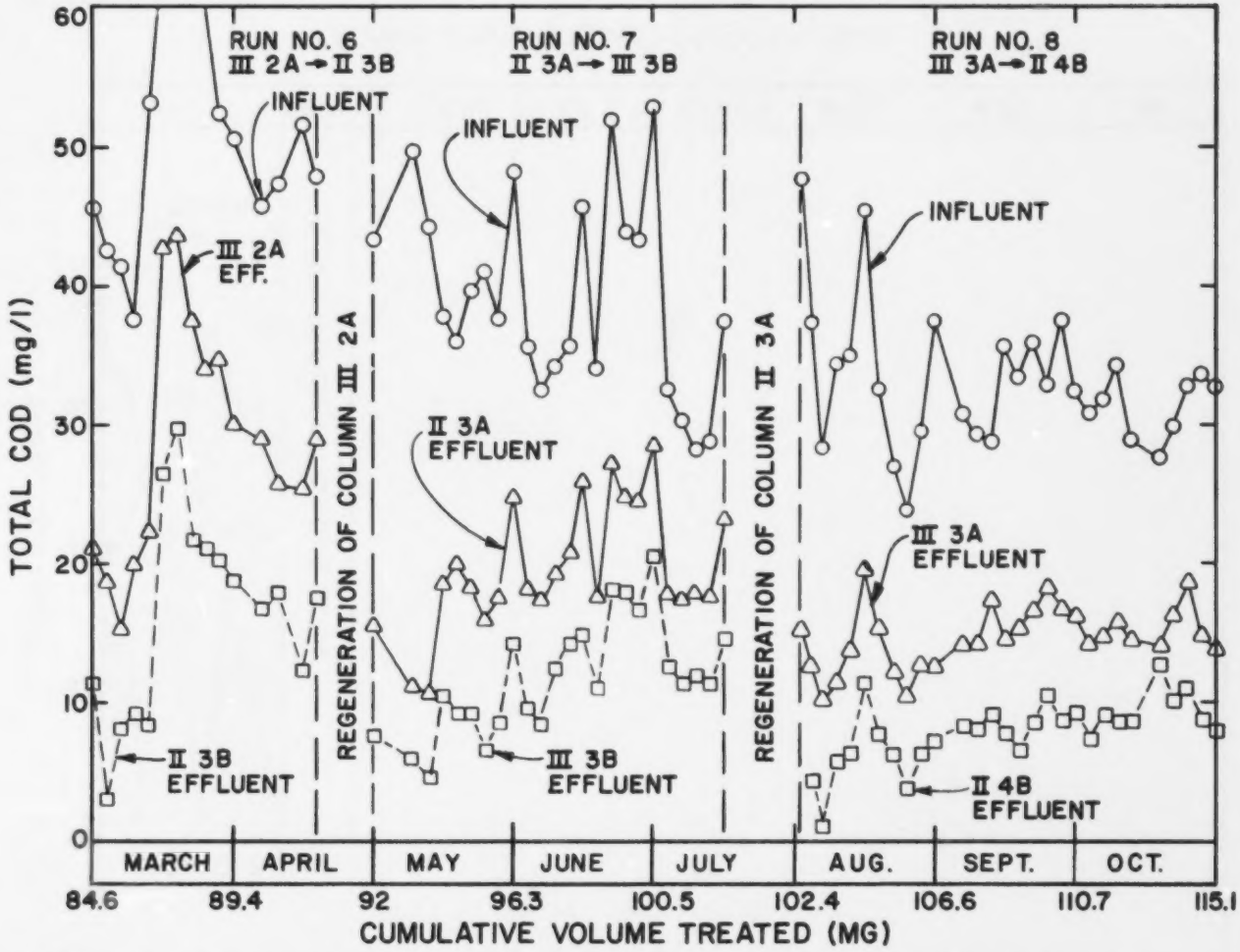

FIGURE 18: COD REMOVAL IN THE FILTRASORB 300 SYSTEM (RUN 6-8)

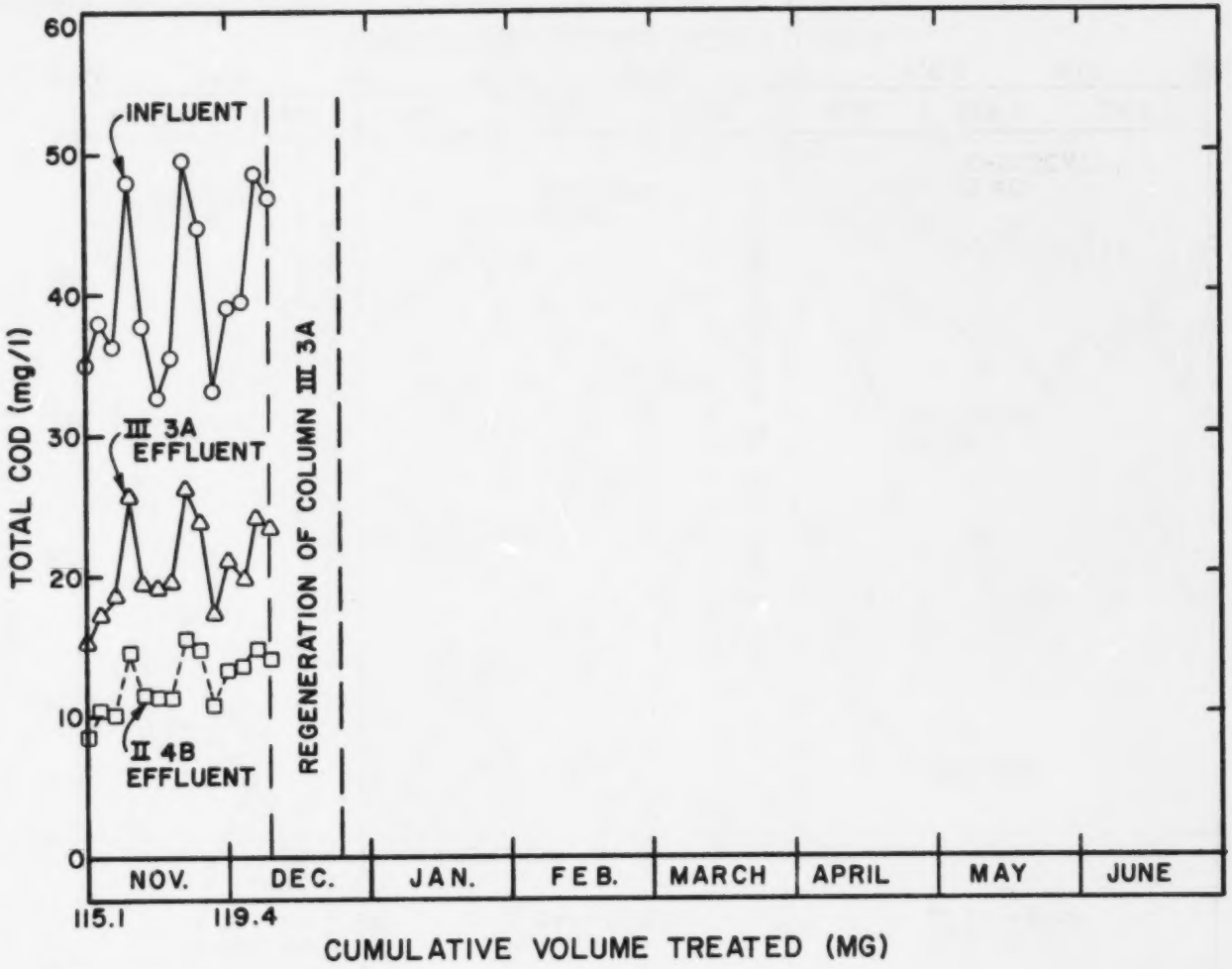

FIGURE 18: CONTINUED

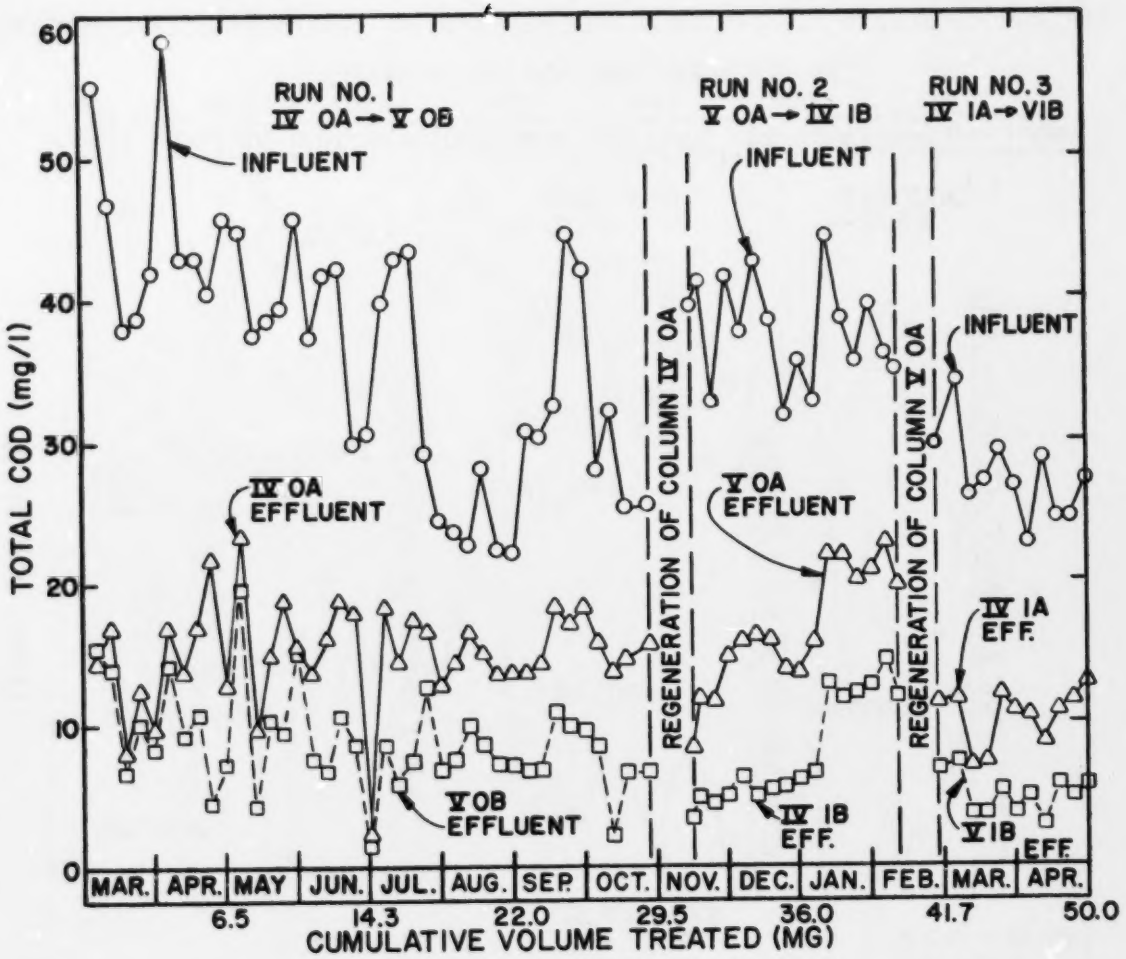

FIGURE 19: COD REMOVAL IN THE FILTRASORB 400 SYSTEM (Run 1-3)

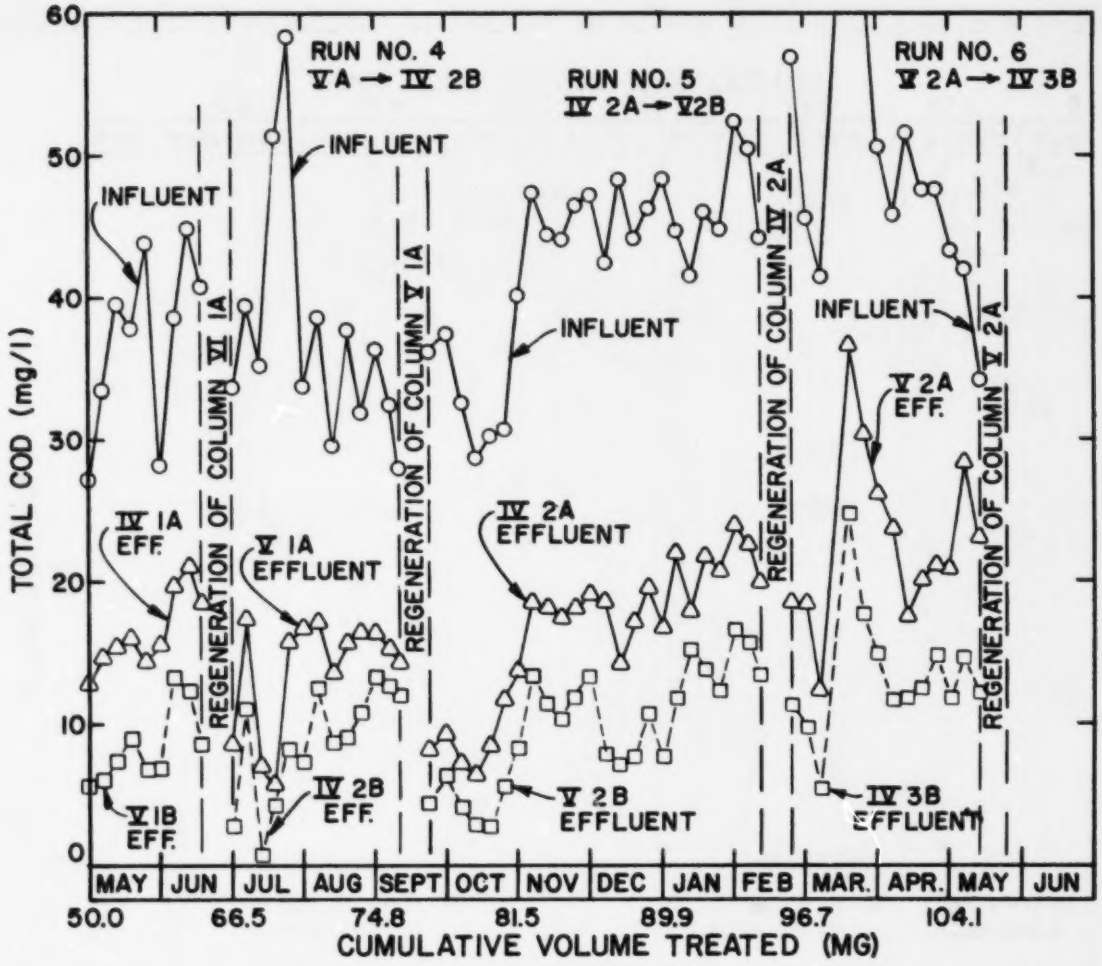

FIGURE 20: COD REMOVAL IN THE FILTRASORB 400 SYSTEM (Run 4-6)

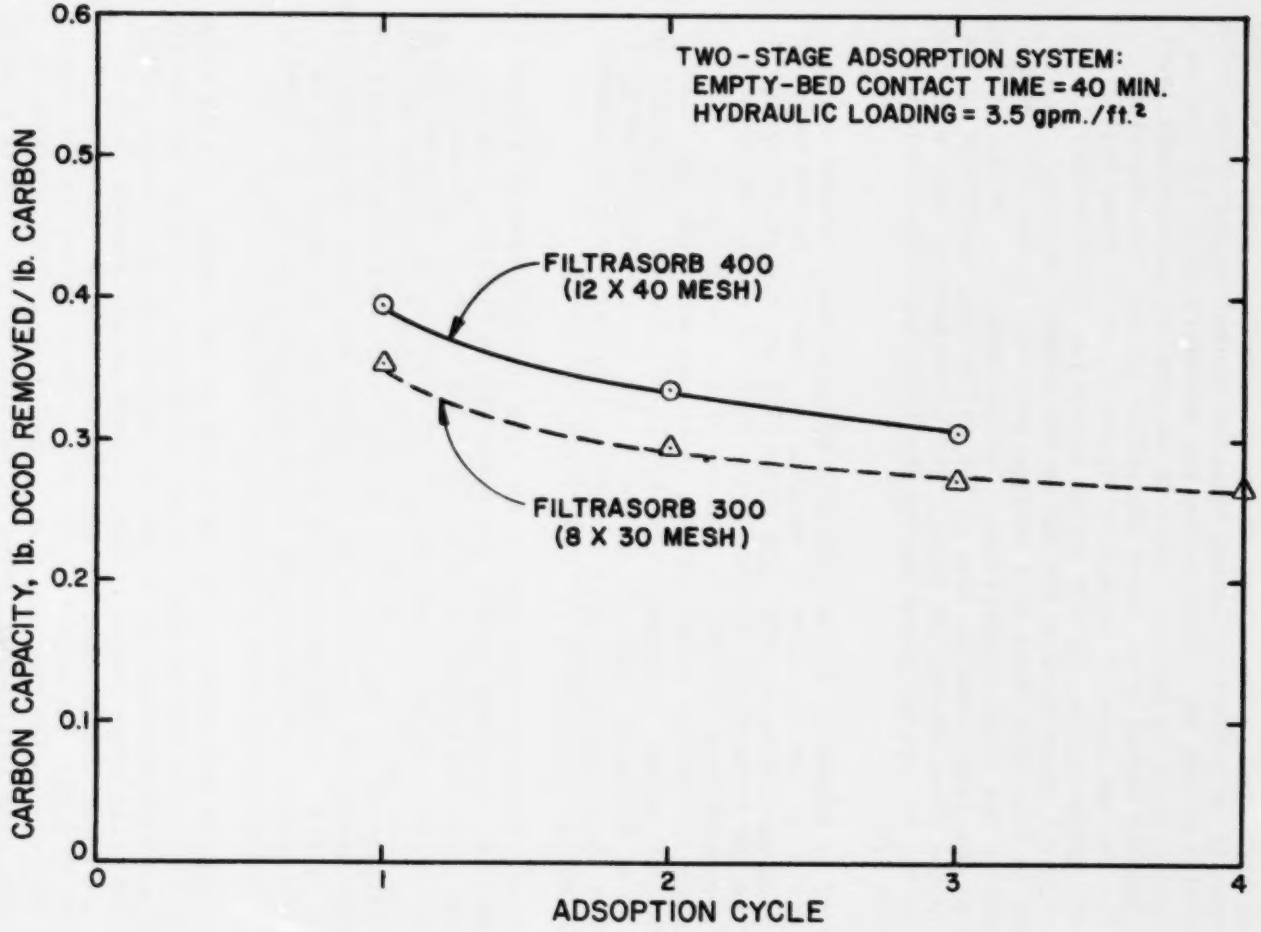

FIGURE 21: EFFECT OF ADSOPTION CYCLE ON CARBON CAPACITY

four adsorption cycles. The decrease in capacity, however, tends to taper off during the last two adsorption cycles. The data indicate that the carbon capacity for DCOD appears to reach a steady-state level of about 0.26 kg DCOD removed/kg carbon after the third adsorption sequence. In Figure 21 is also plotted the carbon capacity for the Filtrasorb 400 system. As indicated in the figure, the same decreasing trend in the DCOD removal capacity is evident following repeated thermal regenerations. The carbon capacity of the 400 system carbon has decreased about 23% from an initial level of 0.39 kg DCOD/kg carbon to an apparent steady-state level of 0.3 kg DCOD/kg carbon after three adsorption cycles. In comparing the data for the two types of carbon in Figure 21, it is shown that at the same adsorption cycle, the 400 system carbon has approximately 13% more DCOD removal capacity than the 300 system carbon. Moreover, it is of interest to note that there is a considerable loss in the DCOD removal capacity during the first regeneration. For instance, for the 300 system, about 69% of the total capacity loss in the entire four adsorption cycles occurred in the first adsorption cycle. The corresponding loss in carbon capacity in the 400 system carbon during the first adsorption cycle was estimated at 67%.

In Figure 22 is presented the relationship between carbon dosage and adsorption cycle for the two types of carbon. For the 300 system carbon, the carbon dosage increased from .038 kg/m^3 (320 lbs carbon/million gallons) during the first cycle to a projected level of about .066 kg/m^3 (550 lbs carbon/million gallons) at the end of the fourth cycle. The carbon dosage for the 400 system carbon, which was about 10% less than that of the 300 system, increased from .034 kg/m^3 (280 lbs carbon/million gallons) to an estimated level of .06 kg/m^3 (500 lbs carbon/million gallons), after three adsorption cycles. Table 10 presents the calculation for carbon dosage and capacity for Filtrasorb 300 carbon.

The effect of carbon particle size on the TCOD and DCOD removal through the two parallel two-stage systems are presented in Figures 23 and 24. The plotted points were obtained by dividing the amount of COD applied and removed by each system in each adsorption sequence by the total weight of carbon in that system. It is apparent from these figures that the COD removal efficiencies of the two-stage systems remain practically constant through four adsorption cycles in the 300 system and three adsorption cycles in the 400 system. Furthermore, while both systems are about equal in terms of COD removal efficiency, the 400 system carbon shows consistently a slightly better performance compared to the 300 system carbon.

In evaluating the data relating to the degree of COD removal in each stage of the parallel two-stage systems, it was found that the major portion of both TCOD and DCOD were removed in the "A" position or first stage carbon column. The first stage carbon columns removed on the average 76.4% of the TCOD and 64.3% DCOD whereas the second stage columns removed only 23.5% TCOD and 35.7% DCOD. The significantly higher TCOD removal in the first stage is consistent with the fact that in two-stage systems operated on downflow mode, the first stage carbon column acts as an effective granular filter bed for the removal of suspended solids. Moreover, the observed low DCOD removal capacity in the second stage, which amounted to about 56% of that in the first stage, could be attributed to the lower DCOD loading applied to the second stage column. Table 11 presents the summary data on the

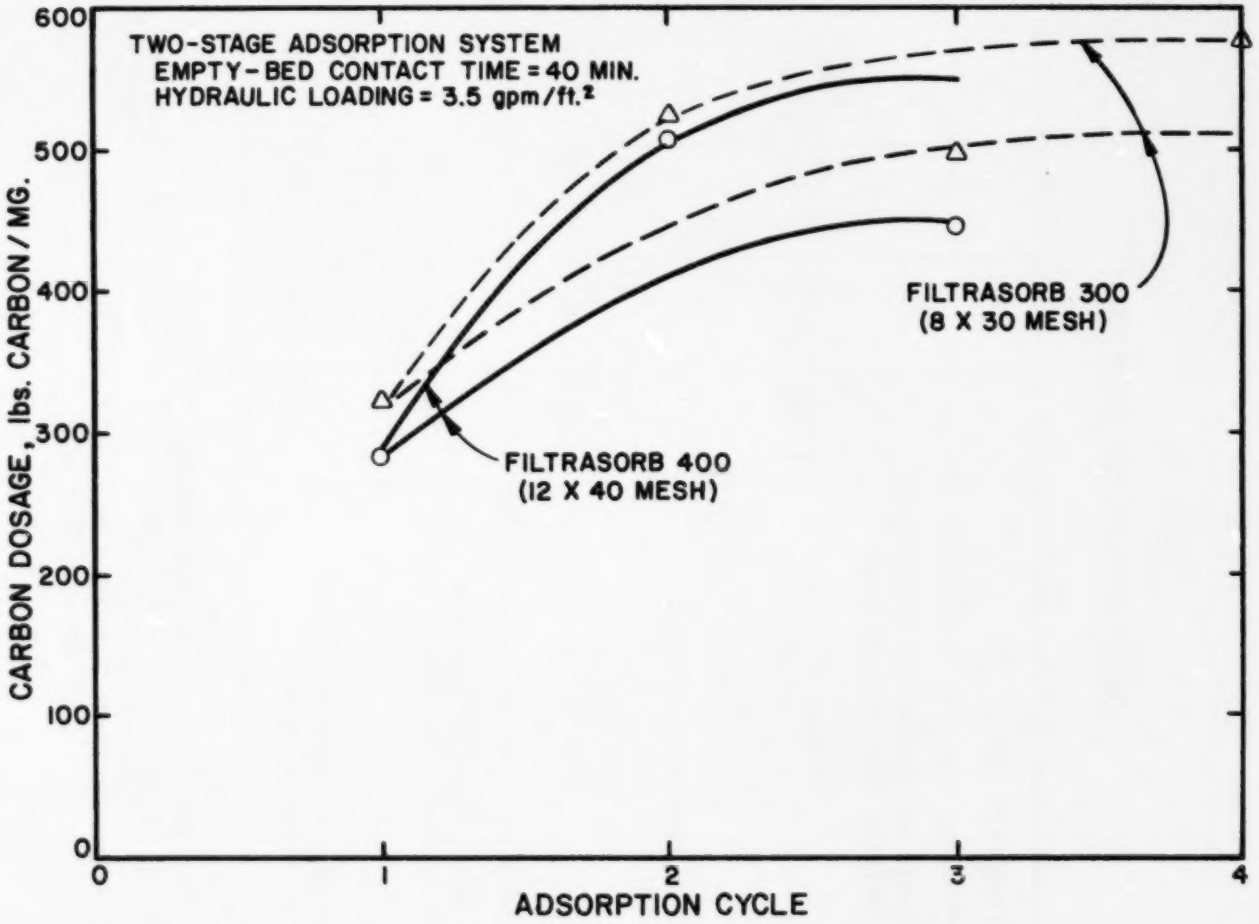

FIGURE 22: EFFECT OF ADSORPTION CYCLE ON CARBON DOSAGE

TABLE 10

CARBON CAPACITY AND DOSAGE DATA

Regeneration Level	0	1	2	3
1. Total Volume Treated in A and B Position, (million gallons)				
Column II	26.324	28.594	27.550	19.830
Column III	38.564	24.609	28.714	28.484
Total	64.888	53.203	56.264	48.313
2. DCOD Removed in A and B Position (lbs.)				
Column II	2742	2145	2019	1615
Column III	2176	1889	1733	2036
Total	4918	4034	3752	3651
3. Carbon Capacity[a], lb DCOD Removed per lb. Carbon	0.35	0.29	0.27	0.26
4. Carbon Dosage[b] lbs carbon per million gallons	322	523	495	576

6960 lbs. carbon per column

a Carbon capacity = Total DCOD removed ÷ 2(6960)

b Carbon dosage = (Total DCOD Removed) 2* ÷ (Total Volume x Carbon capacity)

* 1.5 for regeneration level "0"

Unit conversions: lb x 0.454 = kg

mil gal x 3785 = cu m

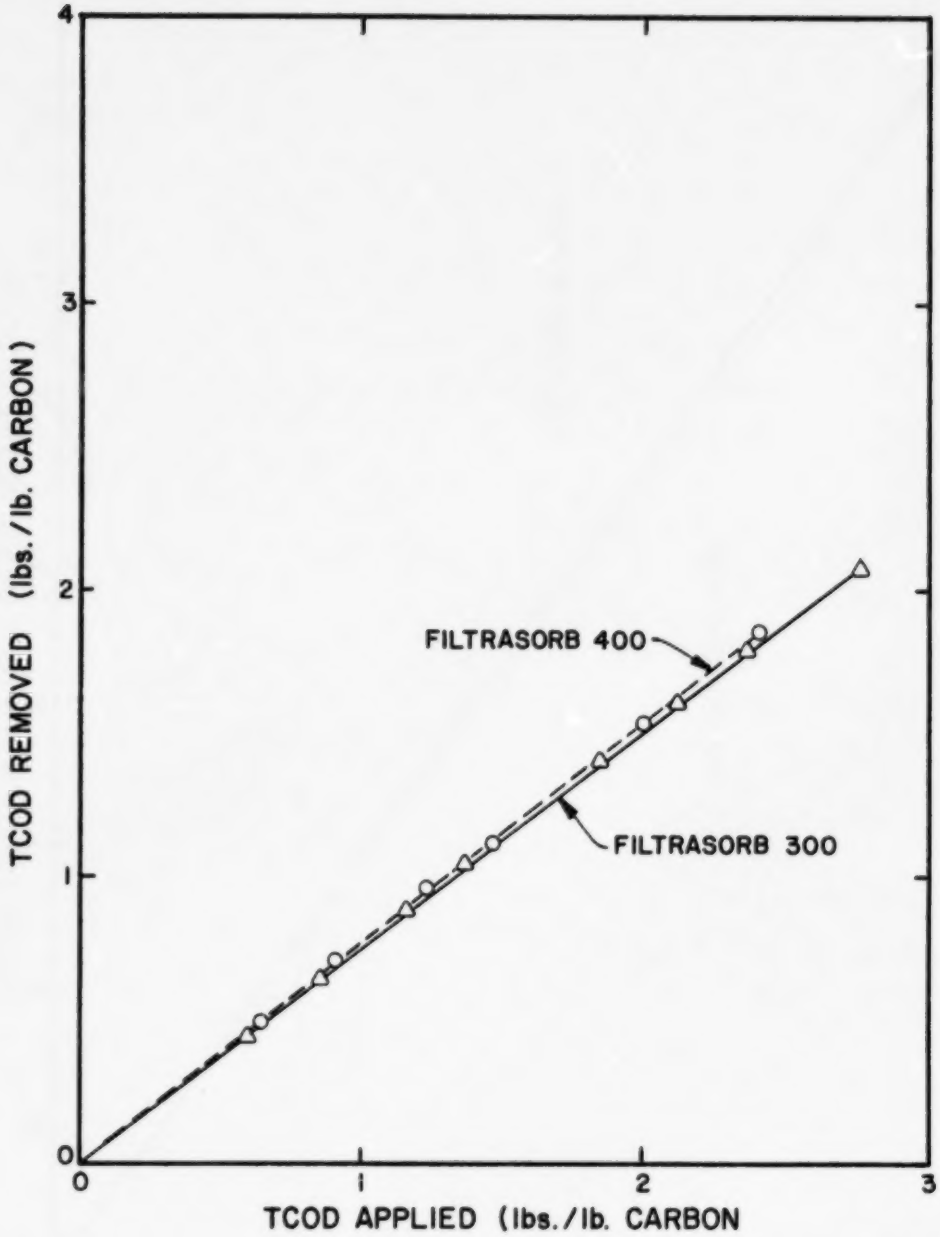

FIGURE 23: EFFECT OF CARBON PARTICLE SIZE ON TCOD REMOVAL

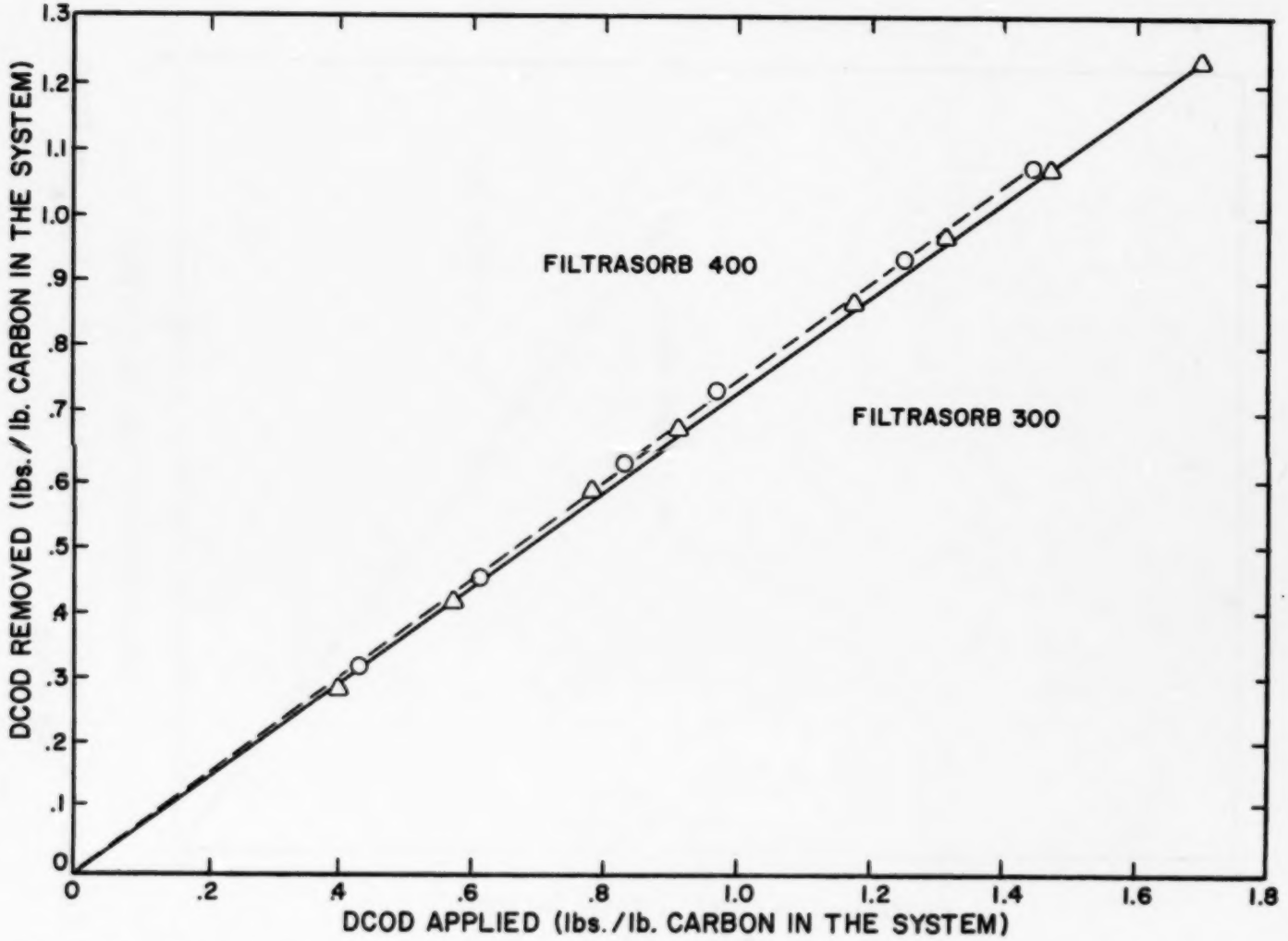

FIGURE 24: EFFECT OF CARBON PARTICLE SIZE ON DCOD REMOVAL

TABLE 11

DISTRIBUTION OF COD REMOVAL IN THE
TWO-STAGE SYSTEMS

Column Position	Filtrasorb 300 System				Filtrasorb 400 System			
	TCOD		DCOD		TCOD		DCOD	
	% Removed	lb removed / lb. c	% Removed	lb removed / lb. c	% Removed	lb removed / lb. c	% Removed	lb removed / lb. c
"A" (1st Stage)	76.4	0.394	64.3	0.180	76.5	0.482	64.3	0.238
"B" (2nd Stage)	23.6	0.122	35.7	0.100	23.5	0.148	35.7	0.132
Total	100	0.516	100	0.28	100	0.630	100	0.370

Unit Conversion: lb x .454 = kg

average COD removal efficiency and carbon capacity in each stage of the two-stage systems.

Effluent Water Quality

In Table 12 are presented the average water quality characteristics of the influent and effluent from the two-stage systems. The data from this table shows that the two-stage systems were effective in producing carbon effluents with low levels of organic content, as measured by both COD and TOC. The TCOD removal has averaged 74.7% in the 300 system and 78.6% in the 400 system. It is further evident from the tabulated data that the 400 system was slightly more effective than the 300 system in the reduction of all the process parameters evaluated. The performance of the two-stage adsorption systems in each adsorption sequence is summarized in Table 13 for the Filtrasorb 300 system and Table 14 for the Filtrasorb 400 system.

The secondary effluent used during the first two adsorption cycles of the two-stage systems was partially nitrified with an average nitrate concentration of 3.7 mg/l N. Thereafter, the average nitrate concentration of the carbon influent was only 1.2 mg/l N as a result of a change in the mode of operation of the activated sludge plant. While the level of nitrate in the carbon influent had no apparent effect on the overall carbon effluent quality, one carbon effluent quality characteristics of significance is the sulfide concentration. Towards the end of the third adsorption cycle when the carbon column influent nitrate concentration averaged only 1.0 mg/l N, hydrogen sulfide odor became evident in the carbon effluent. Results of sulfide determination, which was not performed until the last few days of the third adsorption cycle, indicated averaged total sulfide levels in the carbon effluent of 0.2 mg/l S in the Filtrasorb 300 system and 0.5 mg/l S in the Filtrasorb 400 system. In previous carbon column operations in which the column influent nitrate concentration was about 4 mg/l N or greater, no discernible H_2S odor was detected in the carbon effluent. This absence of detectable H_2S odor in the carbon effluent was attributed to the presence of sufficient amount of influent nitrate concentration which inhibited sulfate reduction within the carbon columns.

During the course of the study, virus analysis was performed on the two-stage system effluent. The results of the virus analysis, without any distinction as to the type of carbon used for the two-stage system, indicated that about 39.4% of the 132 samples analyzed were positive for virus. A summary of the virus testing program for the two-stage system, as well as the other carbon contacting systems, are presented in Table 15. It is evident from the data presented that while granular activated carbon columns effect some degree of virus removal, they cannot be relied upon for consistent virus removal.

COST ESTIMATE

The economic analysis is based on the carbon treatment of the Pomona activated sludge plant effluent for an average design flow of 0.44 m^3/sec

TABLE 12

AVERAGE WATER QUALITY CHARACTERISTICS

Parameters	Filtrasorb 300 (8 x 30 mesh)			Filtrasorb 400 (12 x 40 mesh)		
	Influent	Effluent	Removal, %	Influent	Effluent	Removal, %
Suspended Solids, mg/l	12.2	2.1	82.9	12	2	83.3
Total COD, mg/l	38.7	9.8	74.7	40.2	8.6	78.6
Dissolved COD, mg/l	23.7	6.4	73	24.3	6.2	74.5
TOC, mg/l	12.3	4.0	67.5	10.8	2.6	75.9
Dissolved TOC, mg/l	9.0	3.2	64.4	7.8	1.8	76.9
Turbidity, JTU	7.6	1.8	76.3	7.3	1.4	80.8
MBAS, mg/l	0.36	.04	88.9	.24	.04	83.3
Nitrate-N, mg/l	2.7	1.2	55.6	3.3	1.4	57.6
Color	34.6	5.8	83.2	36.2	4.6	87.3
Duration of run (months)	33			26		
Total volume treated, million gallons	120.53			97.61		

Unit conversions: mil gal x 3785 = cu m

TABLE 13

PERFORMANCE OF THE FILTRASORB 300 TWO-STAGE ADSORPTION SYSTEM

Ads. Cycle	Run No.	Operating Sequence	Two-Stage System Removal, %								
			Total COD	Diss. COD	Total TOC	Diss. TOC	MBAS	COLOR	TURB	NH_3-N	NO_3-N
1	1	II0A III0B	75.8	71.2	77.7	76.5	85.8	82.8	72.2	1.1	41.6
	2	II0A III1B	78.6	73.4	71.8	67.6	87.6	68.6	74.3	3.2	76.9
2	3	III1A III1B	75.6	74.6	76.6	76.8	88.5	89.2	73.8	3.0	8.1
	4	III1A II2B	75.6	73.9	81.4	78.9	89.7	88.1	72.0	3.6	81
3	5	II2A III2B	75.2	69.9	71.1	70.5	84.3	83.5	79.0	0	57.4
	6	III2A II3B	72.6	67.3	62.6	50.5	93.3	79.2	73.8	0	68.6
4	7	II3A III3B	70.3	70.7	38.6	35.2	77.4	80.2	64.3	0	50.6
	8	III3A II4B	73	67.2	-	-	78.7	91.2	81.5	0	80.1

TABLE 14

PERFORMANCE OF THE FILTRASORB 400 TWO-STAGE ADSORPTION SYSTEM

Ads. Cycle	Run No.	Operating Sequence	Two-Stage System Removal, %								
			Total COD	Diss. COD	Total TOC	Diss. TOC	MBAS	COLOR	TURB.	NH_3-N	NO_3-N
1	1	IV0A V0B	76	72.4	75.2	76.8	84.8	86.2	78.9	1.6	36
	2	V0A IV1B	77.5	72.5	67.9	70.9	90.7	82.1	76.0	4.7	73.5
2	3	IV1A V1B	80.0	79.8	82.5	84.8	89.7	91.7	77.9	5.6	89.8
	4	V1A IV2B	75.6	74.2	82.4	79.8	89.9	91.0	73.0	8.6	88.4
3	5	IV2A V2B	78.6	72.9	75.2	75.6	82.3	86.8	81.2	0	59.2
	6	V2A IV3B	74.4	70.5	66.2	65.5	-	87.9	75.5	1.8	77.0

TABLE 15

VIRUS REMOVAL DATA

Sample Type	Sampling Period	Contact Time Minutes	Hydraulic Loading gpm/ft^2	No. of Samples Tested	% of Samples Positive
Activated Sludge Plant Effluent (unchlorinated)	8/7/69 - 3/5/69			28	82.1
Single-stage Carbon Effluent (Filtrasorb 400)	6/5/68 - 1/15/69	10	7	21	52.4
Two-Stage Carbon Effluent (Filtrasorb 400 and 300)	3/18/70 - 11/29/72	40	3.5	132	39.4
Four-stage Carbon Effluent (Filtrasorb 400)	1/15/69 - 3/11/70	40	7.0	59	45.8

Unit conversion: gpm/ft^2 x .0407 = cu m/day/m^2

(10 mgd) and a peak flow of 0.61 m^3/sec (14 mgd). The design parameters for sizing the carbon contactors and the carbon regeneration furnace are presented in Table 16. The table also includes assumed unit costs for carbon and other direct costs for estimating the operation and maintenance (O/M) costs.

In the preparation of the cost estimate, various published reports (2,7,8,9, 10,11) were consulted. The capital costs obtained from literature were adjusted to the EPA sewage treatment plant construction cost index of 185.

The carbon contacting system consists of six trains of two-stage, packed-bed, downflow carbon columns designed for a hydraulic surface loading of 4.75 l/sec/m^2 (7 gpm/ft^2) and an empty-bed contact time of 20 minutes per stage. In order to facilitate uninterrupted operation during carbon regeneration, two spare vessels are provided in addition to the six trains of two-stage carbon columns. One of these spare vessels is initially charged with 36,318 kg (79,900 lbs) of Filtrasorb 300 (8x30 mesh) carbon while the other vessel is reserved for spent carbon storage. Thus, the initial carbon charge is equivalent to the effective volume of 13 carbon contactors.

Table 17 presents a summary of the estimated carbon treatment cost. The cost breakdown shows that capital amortization represents about 58% of the total treatment cost of 11.52¢/1000 gallons. Moreover, the cost of make-up carbon and the cost of operating and maintenance labor represent, respectively, 35.5% and 40.4% of the total O/M costs.

TABLE 16

DESIGN CRITERIA AND UNIT COSTS FOR ECONOMIC ANALYSIS

CARBON CONTACTING SYSTEM	
Empty-bed contact time, min/stage	20
Hydraulic surface loading, gpm/ft^2	7
Backwash volume, % of plant flow	2
Carbon dosage, lb C/MG	550
Carbon Regeneration loss, %	7
CARBON COSTS	
Filtrasorb 300 (8x30 mesh) $/lb	0.40
OPERATING COSTS	
Power, ¢/KWH	2
Fuel, ¢/Therm.	8
Carbon Regeneration Fuel consumption, BTU/lbc	6,000
Backwash Water, ¢/1000 gallons	3
Operating Labor, 4 at $12,000/yr	48,000
Maintenance Labor, 1 at $10,000/yr	10,000
Laboratory Personnel, 1 at $14,000/yr	14,000
Maintenance Materials, $	5,000
CAPITAL COSTS	
All equipment costs were amortized at 6% for 25 years.	

TABLE 17

ESTIMATED GRANULAR ACTIVATED CARBON TREATMENT COSTS*

(10 MGD)

CAPITAL COSTS (1000 of $)	
Influent Pumping	175
Initial Carbon Charge	415
Carbon Contacting System	2,100
Carbon Regeneration System	410
Total Capital Cost	3,100
Amortized Cost (¢/1000 gallons)	6.64
OPERATING AND MAINTENANCE COSTS (¢/1000 gallons)	
Carbon Make-up	1.73
Backwash Water	0.06
Power	0.72
Fuel	0.26
Operating and Maintenance Labor	1.97
Maintenance Materials	0.14
Total O/M Costs	4.88
Total Treatment Cost, (¢/1000 gallons)	11.52

* Based on estimated EPA sewage treatment construction cost index of 185 - October 1973.

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